REFINER
AND NATURAL GASOLINE MANUFACTURER

MONO-CAST
DOUBLEX SIMPLEX CAST IRON PIPE

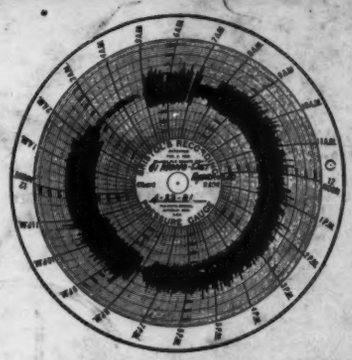
is now in service on 400 pounds natural gas transmission and up to 900 pounds in other services!

Mono-Cast Doublex Simplex Cast Iron Pipe being installed in high pressure (400-lbs.) natural gas transmission line.

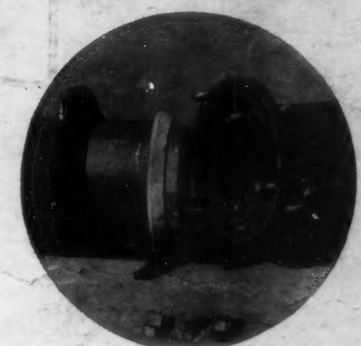
THIS cast iron pipe is made by the Mono-Cast centrifugal method and has a gland-packed joint. The superior quality of this product—its high strength—enables it to meet all demands for practical pipe line construction; strength in light weights, speed and economy in laying, and leakproofness under conditions demanding deflection, longitudinal expansion and contraction, and vibration. It is economical in first cost, and from the standpoint of resistance to corrosion it is in a field by itself. Its resistance is native—it doesn't require wrapping, coating or other artificial protection. Under widely varying conditions, Mono-Cast Doublex Simplex Cast Iron Pipe has definitely proven itself entirely satisfactory for use in high pressure gas and oil lines. Let us cite you this proof and quote you on your pipe requirements. Write nearest Acipco office for a complimentary copy of our catalog: "Mono-Cast Doublex Simplex Cast Iron Pipe."

AMERICAN CAST IRON PIPE COMPANY
BIRMINGHAM, ALABAMA

SAN FRANCISCO	CHICAGO	KANSAS CITY
DALLAS	MINNEAPOLIS	LOS ANGELES
NEW YORK CITY	SEATTLE	CLEVELAND



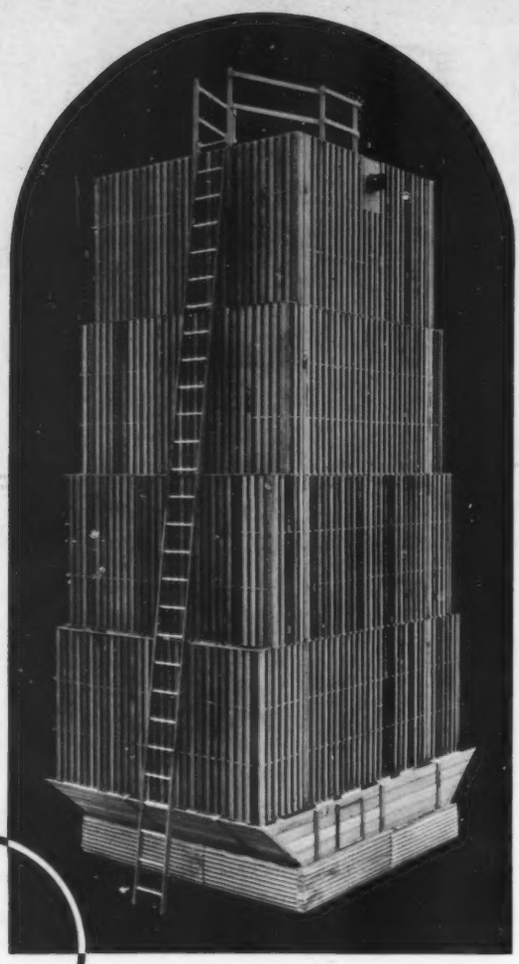
Test Chart showing 900 pounds pressure borne by Mono-Cast Doublex Simplex Cast Iron Pipe in a hydraulic line.



The DOUBLEX SIMPLEX JOINT

When the plain end of a Doublex Simplex pipe is inserted in the socket, a triangular stuffing box is formed with a shoulder against which a gasket is compressed by the gland as the bolts are tightened. A very definite "seat" inside the socket acts to confine the gasket, resulting in maximum compression area which grips the surfaces of the pipe and makes a bottle-tight joint, while the large cross-section of the gasket provides for ample deflection and longitudinal expansion and contraction in the line. Plain, duck-tipped, and lead-tipped rubber, also asbestos gaskets are available. Doublex Simplex Pipe is so simple in its construction that it can be easily and correctly laid with unskilled labor. Only one tool—ratchet wrenches—is necessary to make up the joints.

132R



Presenting the Latest FLUOR Development in Cooling Towers

This brilliant designing and engineering achievement by Fluor engineers marks an advanced phase, a new era in cooling tower manufacture. The tower pictured above has been subjected to severe atmospheric tests under operating conditions. Even those who perfected the unit are amazed at the success of repeated performances. ¶ Increased conservation of water and economy in operation are the fundamentals of this new cooling tower.

It achieves the following points in efficiency and saving, which modern industry will quickly recognize and adopt: ¶ Water loss hardly perceptible. Greater cooling performance. Relatively greater efficiency during periods of no wind, accountable to natural draft effect . . . a feature of construction. Lower cost. Swifter erection. ¶ All requests for information on this successful new tower are being promptly answered.

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Refining in 1932

By GEORGE REID
Associate Editor

SEVERAL factors exist which necessitate continued expansion of crude oil processing facilities in the refining industry. Chief among them is the present lack of balance between facilities for skimming crude for the production of relatively low octane number, straight run gasoline and cracking equipment needed for the production of high octane number motor fuel.

With the present unbalanced condition existing in several refining districts it is necessary to increase the production of high anti-knock value gasoline through the early installation of lower liquid and vapor phase cracking units and gasoline reforming systems, or through the use of cracking coils designed for cracking crude oil direct. Whether in answer to the motorist's demands for higher octane rating, or because of competitive conditions in the marketing branch of the industry, or because of anticipation of the needs of motors of next year or thereafter, the refining industry is bent upon distributing premium and regular grades of gasoline of better quality than standards of last year.

Another factor in addition to the sound or unsound competitive raising of octane ratings of motor fuels which contributes to the necessity for cracking expansion is the decreasing demand for fuel oil by industrial consumers. With a fuel oil market glutted it is far better practice to produce more gasoline by cracking gas oil and fuel oil than to increase fuel oil stocks through the distillation of crude oil, while producing a straightrun gasoline so low in octane rating that it must then be cracked or reformed to meet current popular demand.

It is apparent that the refining industry will process perhaps less crude in 1932 than in 1931, and at the same time increase the total yield of gasoline slightly through the use of additional cracking facilities. It is thought that the industry could well use an average of 60 per cent cracked gasoline to 40 per cent straightrun, which proportions would assure finished products of sufficiently high octane rating to meet current needs for both regular and ethylized or premium grades of gasoline. Such a situation demands the installation of facilities designed for production of high octane rating motor fuels. Gasoline production is now at the approximate proportions of 25 per cent straightrun and 21 per cent cracked gasoline based on the crude charged to stills. During the past few years the proportion of cracked motor fuel has been steadily increasing and the time is not far distant when more motor fuel will be produced through the cracking facilities of the industry than is produced by straight distillation processing.

REPLACEMENT

Because of constant technical improvement in the cracking field, and because of the severity of the service, obsolescence and "old age" become very important factors in both liquid and vapor phase cracking work. The same is true to a somewhat lesser degree in regard to atmospheric and vacuum distillation units of the modern tubular type.

At the beginning of 1931 there were 825 shell still type cracking units reported in this country, all of which, in the light of present day progress may be classed as obsolete, uneconomical and inefficient in their operation; and for the most part they are unsafe. Their total reported capacity at January, 1931, was 171,921 barrels per day. About one-third have been dismantled during the year. The present capacity of this type of obsolete cracking equipment is estimated at 115,000 barrels. Replacement of this old equipment with modern cracking facilities will involve the erection of 35 to 40 cracking units (basing

this on average capacity of 3000 barrels per unit). Approximately half of this replacable capacity is located at the refineries of the 20 major companies operating in the United States.

In addition, many of the older cracking units installed and operating at the time of the first survey of the Bureau of Mines, 1925, are now obsolete and unsafe for further operation, and work now underway involves dismantling and replacing this worn out equipment. The early 500-barrel units are now being torn out entirely or are being converted to other types of work, and being replaced with modern units ranging in capacity from 3000 barrels to as high as 10,000 barrels per day.

The trends in development work involve practically every important department in the complete refinery, especially those of distillation and cracking as discussed above, and to a marked degree the departments of lubricating oil manufacture and chemical treating. The continued transition from the adoption of one improvement and another involves the expenditure of large sums of money. Maintenance alone, at refineries in this country is estimated to average from two to four cents per barrel of crude run, depending upon the type of operation and the severity of corrosive crudes and distillates. It is estimated that the industry spends $36,000,000 annually for maintenance alone. Its expenditures for new equipment such as cracking units, atmospheric and vacuum distillation units, refrigeration and other lubricating oil processing equipment gasoline recovery and stabilization plants, exchangers, fractionators, and other auxiliary equipment and materials will aggregate much more than the cost of refinery maintenance.

Construction programs engaged in by both large and small refineries can not be attributed to economic necessity inasmuch as present refinery capacity in this country is approximately twice the rating required for the production of gasoline needs of the country and exports. Expansion work, however, is essential because of technical development rendering obsolete many prevailing practices. Construction and expansion work now under way in practically all refining districts is therefore termed "technical expansion."

The changing industrial picture is ever bringing about influences which affect not only the equipment, methods and processes upon which production depends, but which contribute to a major extent the expenditures of industry necessitated by ever approaching obsolescence of machinery and products. With such changes comes progress, and it follows that with progress comes obsolescence; and the alert concerns of industry must follow developments closely which may contribute to profit or loss, or to the changing industrial outlook.

Those familiar with the refining of petroleum know that in this industry progress is fast, and the influences of such swift progress must be anticipated where possible, and such influences of technical and engineering development work in both manufacturing and sales extend into all departments of the refinery; for in every processing step in this industry which produces a multitude of products, new and improved methods are here and more are coming.

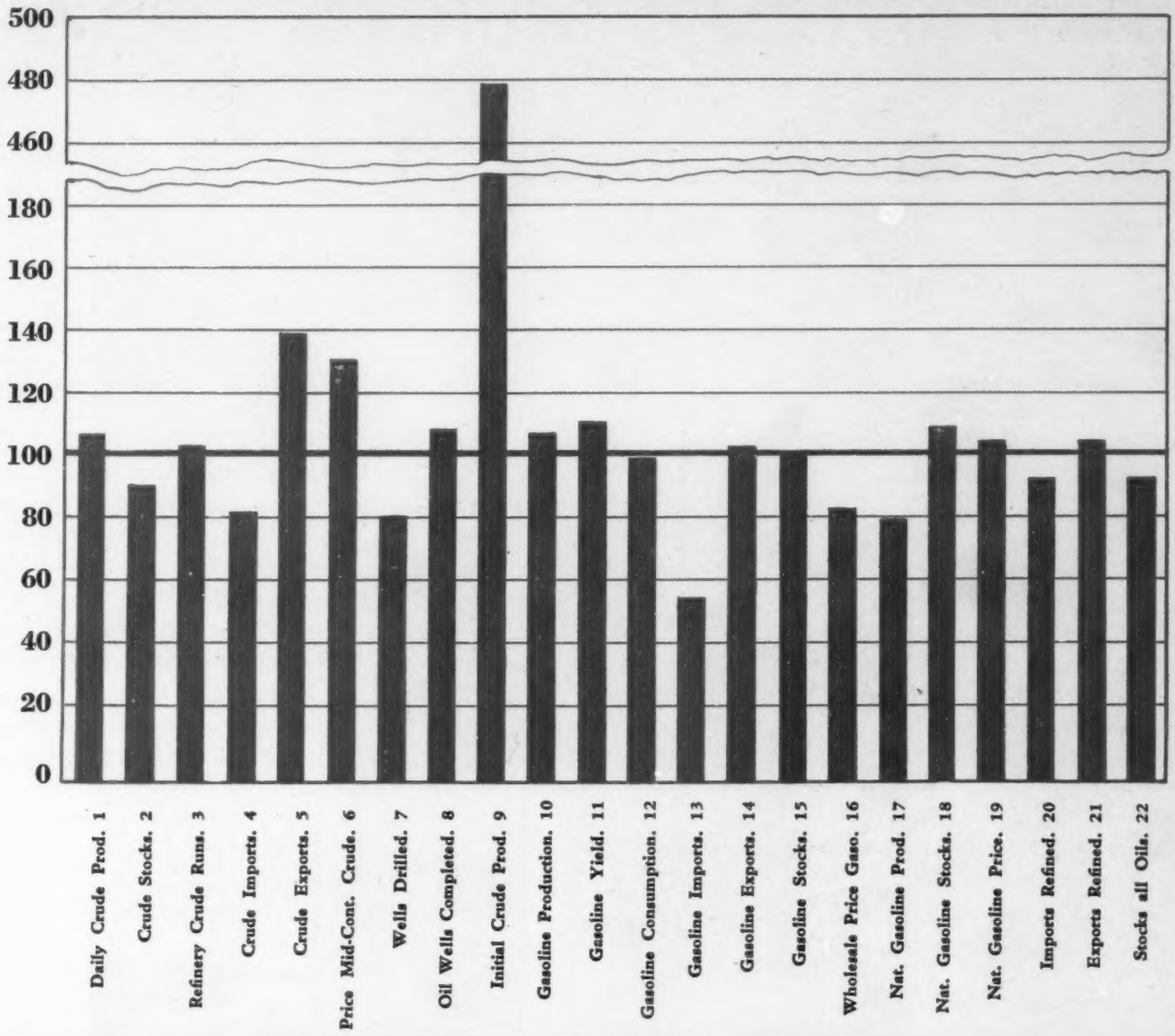
Outside of the cracking field the continued adoption of vapor phase treating methods (two types of processes are now being offered) to avoid excessive volumetric losses and higher costs where sulphuric acid is required in relatively large proportions, is a development brought about by both lower prices for and increasing demand for higher octane number grades of gasoline. With vapor phase treating methods being constantly improved and utilized by the industry the trend points to the gradual reduction of sulfuric consumption by the refining industry, perhaps to the point where this industry may eventually not be a major consumer of this chemical. This applies largely to the refining of gasoline and kerosene, but there are similar developments underway in the chemical treatment of lubricating oils, especially in connection with the growing adoption of the sulfur dioxide extraction process and continuous centrifugal treatment with the acid but using smaller proportions effectively.

Asphalt manufacturing methods have also been undergoing changes tending toward improvement in product and economy of processing. The vacuum flash coil now utilized by many asphalt manufacturers is the most important development in this field as this equipment assures control of the specifications of practically any desired product through the close separation of lighter products in asphaltic crudes. Further, with the development cracking equipment and its controlled operation for asphalt and road oil production it is now possible for practically any refinery to produce these materials synthetically and from any type of crude.

While the technical progress of the industry during the past few years has been remarkable, the future developments will bring about many important changes in manufacturing and processing methods. Interesting progress is reported in oil-gas conversion, gas cracking, alcohol production and solvents manufacture and the production of carbon dioxide.

Such developments point to the continued expansion within the refining industry even though it is now provided with more crude processing facilities than are economically justified. It is technical advancement that brings about the periodic revamping of all process industries. Since the advent of the tube still and the bubble tower the refining industry has been constantly improving its technique, not only in the department of distillation, but also in all departments of the complete refinery.

Salient Oil Industry Factors for November, 1931, Expressed in Per Cent of November, 1930. Heavy Line (100) Represents November, 1930



Oil Industry Barometer—November, 1931

Index of Salient Factors—Current and Cumulative

	Factors—	This Month			11 Months Ending Nov. 30		
		Last Year	This Year	Index (1)	Last Year	This Year	Index (1)
1	Daily Average Crude Production (Thous. Bbls.)....	2,265	2,428	107	2,488	2,327	94
2	Total Crude Stocks (Thous. Bbls.).....	517,021	463,079	90	517,021	463,079	90
3	Refinery Crude Runs (Thous. Bbls.).....	70,310	71,367	102	855,866	820,901	96
4	Crude Imports (Thous. Bbls.).....	4,467	3,604	81	57,402	42,935	75
5	Crude Exports (Thous. Bbls.).....	1,765	2,449	139	22,366	24,475	109
6	Posted Price 36 Grav. Mid-Cont. Crude (Per Bbl.) (2)	$.59	$.77	130	$1.22	$.61	50
7	Wells Drilled*	1,253	1,008	80	17,877	10,533	59
8	Oil Wells Completed*	644	698	108	9,862	5,900	60
9	Initial Crude Production (Thous. Bbls.)*.....	901	4,314	479	8,825	28,241	320
10	Gasoline Production (Thous. Bbls.).....	33,616	36,137	107	406,256	401,837	99
11	Gasoline Yield—Average Per Cent of Crude (4).....	41.66	46.00	110	41.98	44.29	106
12	Domestic Gasoline Consumption (Thous. Bbls.)....	30,984	30,640	99	365,587	372,784	102
13	Gasoline Imports (Thous. Bbls.).....	1,079	586	54	15,963	12,828	80
14	Gasoline Exports (Thous. Bbls.).....	3,690	3,743	101	61,188	43,141	71
15	Gasoline Stocks at Refineries (Thous. Bbls.).....	38,705	38,696	100	38,705	38,696	100
16	Refinery Price U. S. Motor Gaso.—Group 3 (Cents) (3)	5⅓	4⅜	82	6⅜	3⅜	53
17	Natural Gasoline Production (Thous. Bbls.).....	4,317	3,405	79	48,360	39,490	82
18	Natural Gasoline Stocks (Thous. Bbls.).....	2,380	2,586	109	2,380	2,586	109
19	Price Grade "A" Natural Gasoline, Group 3 (Cents (3)	4½	4⅝	104	5⅜	3¼	60
20	Imports of All Refined Oils (Thous. Bbls.).....	2,807	2,579	92	40,179	35,929	89
21	Exports of All Refined Oils (Thous. Bbls.).....	7,768	8,068	104	123,066	92,867	75
22	Total Stocks of All Oils (Thous. Bbls.).....	671,874	618,938	92	671,874	618,938	92
	Weighted Index of Supply	.....	.....	125	.....	.....	111
	Weighted Index of Demand.....	.....	.....	109	.....	.....	90
	Weighted Index of Prices.....	.....	.....	105	.....	.....	54

*Exclusive of Rocky Mountains. (1)—Last Year—100.
(2)—Weighted Average. (3)—Monthly Average.
(4)—Straight run and cracked gasoline only.

See Additional Economics in Month in Review Section, Page 51a

A RECENT photograph of the new gasoline recovery plant installed at Continental Oil Company's Ponca City refinery. The first tower on the left is the still and next to it is the debutanizer. The smaller drum just behind this latter piece of equipment is the reflux drum. In the next unit, the first tower is the low pressure absorber. On the right and to the rear are the stabilizer and the high pressure absorber. The cooling tower can be seen in the background to the right.

A New Gasoline Recovery and Stabilization Plant

By J. A. JAMISON
Technologist, Continental Oil Company

SIMULTANEOUSLY with the improvement of the antiknock properties of gasoline, the modern refiner is developing processes to insure the ultimate judge—the consumer—of receiving the benefit of his efforts. Too often the energy expended in providing motor fuel with desirable antiknock characteristics is sacrificed in subsequent acid treatment and evaporation losses.

To insure that the consumer receives the benefit of these improvements, Continental Oil Company installed units for the stabilization of pressure distillate at its Ponca City and Baltimore refineries some time ago. In addition to obtaining a stabilized distillate of negligible propane content (below one-half of one per cent), 10-15 per cent of the light ends were removed and further rectified to provide a suitable light gasoline for blending.

Recently the equipment at the Ponca City plant was revamped and further additions made by Soars & Lovelace, Inc., of Kansas City. The present unit has a stabilizing capacity of 7000 barrels per day of Dubbs distillate; removes the gasoline fractions from 7,250,000 cubic feet of refinery vapors and field gases; and, rectifies the gasoline thus obtained to desired volatility. In addition, the plant can handle at the same time 500 barrels per day of casinghead from outside sources, refractionating the product to desired specifications. A depropanizer unit has also been installed to handle 3000 barrels per day of cracked gasoline from two Cross hot oil units.

The usual stabilization unit, Figure 1, consists of a bubble tower, reflux condenser and reflux drum. The feed enters the tower six to eight plates from the bottom and is fractionated into two portions: the desired stabilized product as bottoms; and, the undesired light ends which leave the reflux drum, as gases. There will usually be found auxiliary equipment in the form of an exchanger to preheat the feed with the hot stabilized bottoms from the unit. To reduce the size of the heating element in the reboiler section, a steam heater may be placed on the feed between the exchanger and the tower. A cooler is provided on the stabilized product from the exchanger. All heating is accomplished in closed coils.

THEORY

The operations connected with a gasoline recovery plant are essentially of a vapor pressure character.[1] The calculation of plant operating conditions such as pressure and temperatures to fulfill a specific duty is based on the concept that equilibrium is established between a liquid and the vapor existing directly above it. Whether the problem is the stabilization of a cracked gasoline or the debutanization of a distillate or the absorption of gasoline from a rich gas, the equilibrium concept furnishes a method of attack.[2] While investigations have shown that equilibrium is not actually complete,[3,4] for the present discussion let us assume that the elementary laws of physical chemistry hold true. As a concrete example, let us consider the liquid in the reflux drum of the still in an absorption system. For the sake of simplicity, assume the liquid to consist of a mixture of propane, n-butane, pentane and hexane at a temperature of t°F. and a pressure on the drum of π lbs. per square inch abs.

The total pressure, π, is the sum of the partial pressures, p, of the individual constituents (use subscripts to denote each individual hydrocarbon).

In the liquid phase, the partial pressure of each constituent is a function of its vapor pressure in the pure state at t°F. and its concentration (expressed as a mol fraction) in the liquid. In other words, Raoult's law states that:

$$P_1 = Px$$

where P_1 = partial pressure exerted by any constituent in the liquid phase, pounds per square inch abs.
P = vapor pressure in the pure state at t°F. of that component, pounds per square inch abs.
x = mol fraction of component in the mixture.

In the vapors existing above the liquid, the partial pressure of any constituent is proportional to the

total pressure and its concentration (mol or volume fraction). Dalton's law states that

$$P_g = \pi y$$

where P_g = partial pressure exerted by any constituent in the vapor phase, pounds per square inch abs.
π = total (plant) pressure, pounds per square inch abs.
y = mol (or volume) fraction of that component in the mixture of vapors.

At equilibrium, the partial pressure exerted in the vapor phase must equal its partial pressure exerted by the liquid phase; that is,

$$P_l = P_g$$
$$Px = \pi y \qquad (O)$$
$$\text{Also, } \pi = p_3 + p_{n_4} + p_5 + p_6 \qquad (A)$$
$$\text{and, } x_3 + x_{n_4} + x_5 + x_6 = 1 \qquad (B)$$

APPLICATION OF EQUATIONS

In the application of the above equations to plant practice, the composition of the unstabilized feed must be known. Likewise the composition of the desired stabilized product will be known. By difference, the composition and amount of the undesirable light ends is obtained.

The condensing equipment available will determine the temperature t°F. possible in the reflux drum. The value of the vapor pressure of the pure components at t°F. is obtained from tables.[5] The mol fraction of each constituent in the vapors (light ends) leaving the reflux drum is known. This leaves two unknowns in Eq. (O); namely, x and π. Eq. (B) furnishes the value of the various x quantities; that is, the concentration of each constituent in the liquid in the reflux drum. Whence, Eqs. (O) and (B) may be rewritten,

$$\frac{\pi y_3}{P_3} + \frac{\pi y_{n_4}}{P_{n_4}} + \frac{\pi y_5}{P_5} + \frac{\pi y_6}{P_6} = 1$$

thus obtaining the absolute operating pressure, π, required.

Having determined the operating pressure, π, it is now necessary to calculate the temperature required in the base of the reboiler.

In this, one must resort to trial and error approximations; substituting the values of P in equation (O) and making the sum of the values of y equal to unity. In other words, the values of π, and x, (concentration of each constituent of the stabilized product) are known; and it is necessary to determine the temperature which will make

$$\frac{P_3 x_3}{\pi} + \frac{P_{n_4} x_{n_4}}{\pi} + \frac{P_5 x_5}{\pi} + \frac{P_6 x_6}{\pi} = 1$$

There is only one temperature which will satisfy the above equation and Equation (B).

Rectification calculations are more easily handled on a mol basis, so that the usual laboratory analysis, expressed on a volume basis must first be converted to a molal basis.

PRESSURE DISTILLATE DEBUTANIZER

The Dubbs stills operate at a plant pressure of 250 pounds ga. The high pressure separator, at 225 pounds separates the gas and distillate produced. The distillate under its own pressure passes to two surge drums made of old reaction chambers, located at the Gasoline Recovery Plant. Any gas which is formed in these surge drums is vented to the high pressure absorber. The unstabilized distillate continues under its own pressure thru a recorder-controller, exchanger, and steam heater to the debutanizer tower. The tower, which contains 24 trays, eight of which are used for stripping, operates at 82 pounds ga. A temperature controller-recorder holds the bottom temperature at 340°F. The overhead at 190°F. passes through a condenser into a combination reflux drum and feed to the final stabilizer. Thus, by raising the temperature at the top of the tower, it is possible to obtain 10 to 12 per cent of light gasoline in excess of the amount required for reflux to the debutanizer. The gases from the reflux drum pass to the low pressure absorber.

The bottoms of the tower, stabilized pressure distillate, pass thru an exchanger against the incoming feed; a cooler and a valve actuated by a liquid level control on the tower.

Typical analyses are

	Unstabilized Distillate	Stabilized Distillate
Ethane	3.18 per cent volume	0.00 per cent by volume
Propane	6.13	0.25
Iso Butane	1.09	0.50
N-Butane	7.82	2.25
Pentane plus	81.87 of 47.5° API	97.00
°A.P.I.		45
I.B.P.		118
E.P.		510
Reid V. T. @ 100°F.		5-6 lbs.

HIGH PRESSURE ABSORBER

The gas from the separators at the Dubbs pressure stills passes directly to the high pressure absorber where an oil rate of 15 gallons per 1,000 standard feet of rich gas is used. A 41° A.P.I. 400° I.B.P. untreated kerosene is used as absorption medium.

The residue gas from the absorber passes through a pressure regulator which maintains a pressure of 190 pounds on the tower. The tower is a 16 plate bubble cap column provided with a mist extractor to prevent oil carry over. The residue gas enters the refinery fuel system.

A centrifugal pump supplies lean oil to the absorber. An oil-gas ratio controller insures the absorber of operating at a constant desired efficiency. The rich oil from the absorber is regulated by a liquid level control.

LOW PRESSURE ABSORBER

The gas charged to the low pressure absorber is made up of any or all of the following in varying proportions:

A. Vacuum gases from producing field south of

the refinery, run tanks of the crude distillation units, and pressure distillate run tanks when units come on stream or shut down.

B. Thirty pounds gas from separators at Cross stills and depropanizer, and Dubbs low pressure separators when units are coming on steam or shutting down.

C. Cycle gases from still and light gasoline stabilizer.

D. Gas from pressure distillate debutanizer.

The vacuum gases are handled by two 90 hp compressors which compress the gases from five inches Hg. to 75 pounds ga. Two more 90 hp compressors receive the 30 pound gases compressing them to 75 pounds ga. The gas from the debutanizer and cycle gases pass under their own pressure directly to the 70 pound separator.

The compressors discharge into a common header and the gases are cooled. Any gasoline condensed is dropped in the 70 pound separator ahead of the low pressure absorber. This low stage gasoline is pumped to the feed drum of the light gasoline stabilizer.

The gas from the separator passes through an oil-gas ratio controller to the absorber. To remove the latent heat of condensation of the gasoline, five sections of water intercoolers were placed in the lower half of the absorber. An oil rate of 53 gals. per M standard feet of rich gas is used when the plant is operated for 90 per cent butane extraction. The lean oil is an untreated kerosene of 41° A.P.I. and 400° I.B.P.

A liquid-level controller is used on the base of the absorber to control the flow of rich oil to the rich oil tank. The vapors from this tank are vented back to the absorber.

The absorber is equipped with 16 bubble cap trays and a mist extractor.

STILL

The rich oil from both the low and high pressure absorbers passes to a rich oil tank, from which a

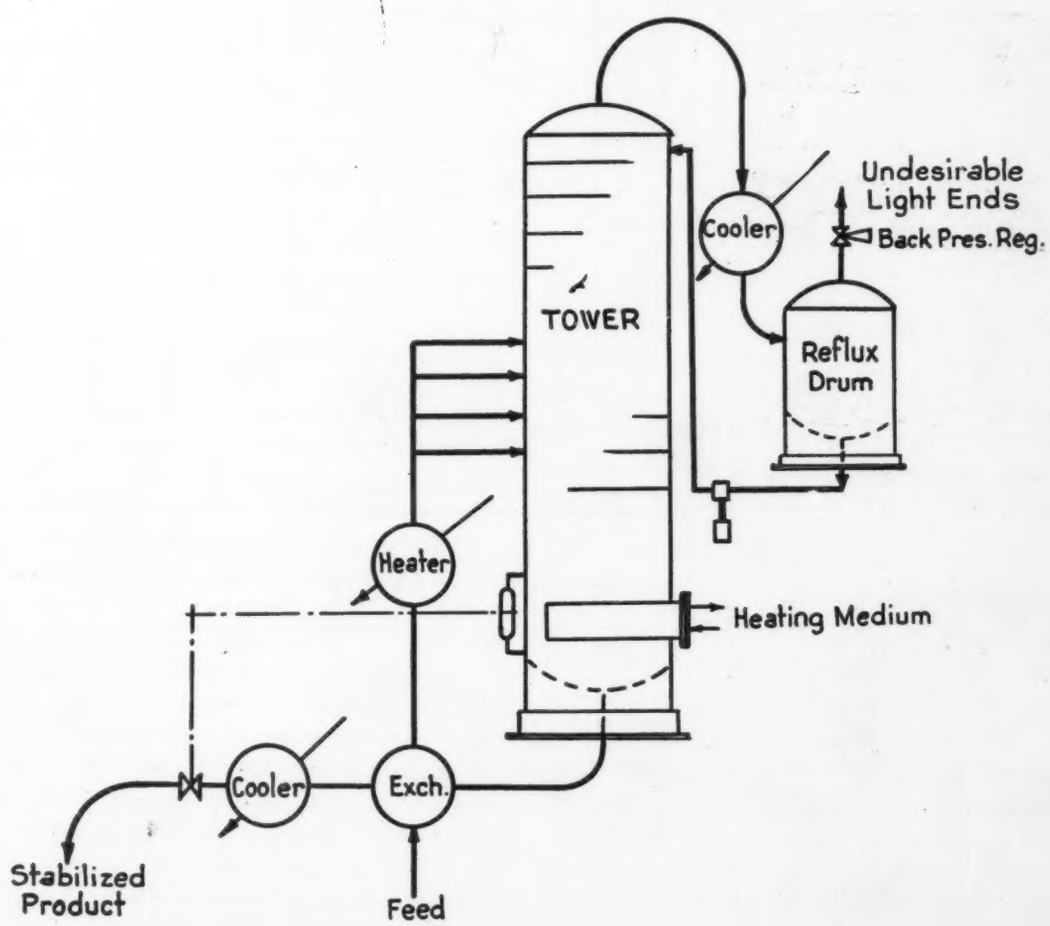
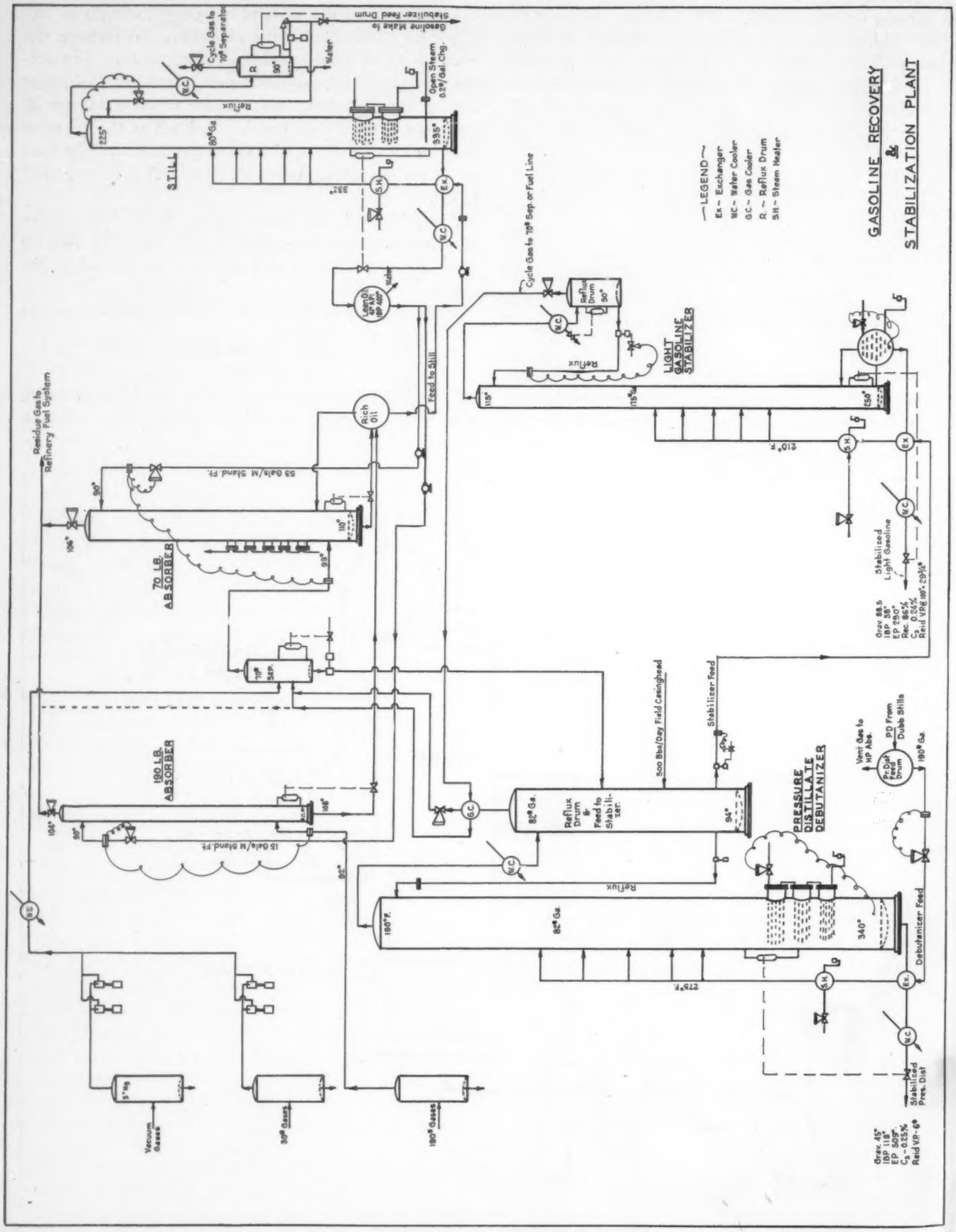


FIG. 1 – STABILIZER UNIT


GASOLINE RECOVERY
&
STABILIZATION PLANT
—LEGEND—
Ex – Exchanger
WC – Water Cooler
GC – Gas Cooler
R – Reflux Drum
SH – Steam Heater
STILL
Cycle Gas to 70# Separator
Gasoline Make to Stabilizer Feed Drum
Water
Reflux
Open Steam 0.2%/Gal Chg.
225°
80# Ga.
332°
335°
Lean Oil 42° API @ 60°F
Cycle Gas to 70# Sep. or Fuel Line
Reflux Drum
90°
LIGHT GASOLINE STABILIZER
Reflux
118°
175# Ga.
259°
210°F.
Feed to Still
Rich Oil
Residue Gas to Refinery Fuel System
53 Gals/M Stand. Ft.
90°
106°
110°
98°
70 LB. ABSORBER
Stabilized Light Gasoline
Grav 88.5
IBP 58°
EP 290°
Rec. 98.5%
C5 – 22.5%
Reid VPR lbs. 29½#
110# SEP.
Stabilizer Feed
500 Bbl/Day Field Casinghead
Vent Gas to HP Abs.
PD From Dubbs Stills
PD Gas Feed Drum
190# Ga.
190 LB. ABSORBER
104°
108°
90°
15 Gals/M Stand. Ft.
Reflux Drum & Feed to Stabilizer
82# Ga.
94°
PRESSURE DISTILLATE DEBUTANIZER
82°
Reflux
180°F.
82# Ga.
340°
Debutanizer Feed
275°F.
5"Hg
Vacuum Gases
30# Gases
180# Gases
Stabilized Pres. Dist.
Grav. 45°
IBP 118°
EP 505°
C5 – 0.15%
Reid VP – 6#

centrifugal pump takes suction. The oil is discharged through a flow indicator to the exchangers, then through a steam heater to the still. The still contains 24 trays, eight of which are used for stripping. Operating at 80 pounds ga. the vent gases from the reflux drum are cycled back to the low pressure absorber. Open and closed steam is used in the still to completely strip the oil of its gasoline content. The water from the condensation of the open steam is removed from both the reflux drum and the lean oil tank. The pressure on the still is sufficient to force the lean oil from the bottom of the tower through the exchangers and water coolers to the lean oil tank which is maintained at 20 pounds gauge.

The gasoline in the reflux drum is picked up by a steam pump which is running continuously. The top temperature controller operates a valve to provide the necessary reflux for the tower. The gasoline-make is operated by a liquid level control from the reflux drum to a valve in the discharge line connected to the light gasoline stabilizer feed drum.

LIGHT GASOLINE STABILIZER

The composite light gasoline from the distillate debutanizer, absorber separator, and still is discharged from its feed drum by a steam pump through a flow controller through the stabilized gasoline exchanger and a steam preheater to the tower. The tower has been made of ample capacity to handle an additional 500 barrels per day of casinghead from outside sources. Heating is completely accomplished by closed steam coils.

Operating for 90 per cent butane extraction, the tower is held at 175 pounds ga. by a pressure regulator on the gases leaving the reflux drum. These cycle gases are expanded in an exchanger thus cooling the gases from the debutanizer reflux drum.

The overhead vapors, at 115°F. pass through a reflux condenser to the reflux drum, which is held at 90°F.

The reboiler temperature of 250°F. is controlled by a thermowell located in the body of the external reboiler. This controller operates a valve on the steam into the reboiler element.

Two methods have been used to control the stabilizer. At first, flow controllers were placed on both the incoming feed and the reflux to the tower, always maintaining a constant ratio by volume. While this method of control would prove excellent for a stabilizer which is receiving a feed stock of constant composition, it is not quite adequate when the quality is changing—for the volumetric reflux ratio is different for each feed stock and with each change in the temperature of the reflux drum.

The second method of control was to fix the temperature and pressure of the reflux drum. Refluxing all of the condensate to the tower automatically compensates for any change in either the rate or composition of the feed. In this way the operator need not know the exact volumetric ratio between the wet reflux and the feed.

The following represents typical products of the light gasoline stabilizer:

	Feed	Stabilized product	Cycle gases
	Per cent by volume	Per cent by volume	Per cent by volume
Ethane	0.31	0.00	18.78
Propane	8.68	0.21	72.05
Iso Butane	4.61	3.38	6.13
N-Butane	29.58	33.80	3.03
Pentane Plus	56.82	62.61	0.00
Grav.	...	88.5	...
Engler Recovery	...	87 per cent	
Reid V. T. @ 100°F.	...	30 lbs.	
Oxygen Bomb	...	4 hr. minimum	

CROSS GASOLINE DEPROPANIZER

Whereas the stabilizer for the light gasoline utilizes steam, the unit at the Cross stills is operated with the waste heat in the fuel oil from the evaporator. The flow is the same as that illustrated in Figure 1 with the elimination of the steam heater.

Operating at 65 pounds ga. pressure and 300° tower base temperature the resulting stabilized gasoline has the following tests:

Grav.	56.5° API
I.B.P.	100
E.P.	410
Propane	0.2 per cent vol.
Reid V. T. @ 100°F.	5-6 pounds
Oxygen Bomb	4 hour minimum

SUMMARY

The stabilization of Dubbs distillate increases the overall gasoline yield for the cracking unit, reduces the volume of distillate to be treated and practically eliminates the evaporation losses on rerunning.

The absorption plant recovers the gasoline from the various refinery and field gases, providing a supply of light gasoline for blending.

The stabilizer provides a means of obtaining any desired volatility light gasoline of negligible propane content. By sweetening in a closed pressure system and storing the completely refined product under pressure, utilizing antiquated Fleming cracking stills for tanks, the refinery is self supporting from the viewpoint of blending requirements.

The depropanizer reduces the propane content and vapor pressure of the Cross gasoline; eliminating tank evaporation loss and providing a perfect motor fuel for all kinds of weather. The gases from this operation may be further processed through the low pressure absorber or by-passed to the refinery fuel system depending on the butane extraction required.

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Control of Gasoline Volatility and Evaporation Losses

By S. D. TURNER and J. W. HARRELL
Humble Oil & Refining Company

WITH the recent installation of a system for the recovery of vapors from finished gasoline storage tanks of 2,500,000 barrels capacity, engineers at the Baytown refinery of the Humble Oil & Refining Company have completed equipment for the handling of volatile products through all stages of refining without evaporation losses. The essential feature of the plan developed for accomplishing this is the segregation from the crude, in its first distillation, or in the case of cracked distillates, from them immediately after their production, of the more highly volatile fractions, leaving the bulk of the naphthas sufficiently stabilized to prevent vapor losses on processing, and the handling of the segregated volatile fractions in specialized equipment, until they are finally blended back into the naphtha to form finished gasoline.

VAPOR RECOVERY AT CRUDE BATTERIES

At the crude distillation batteries, vapors from the naphtha streams are vented, just before reaching the receiving houses, and again at the rundown tanks, to the suction of a compression system. This system, equipped with 14 compressors of 1220 total horsepower, and having a capacity of 6,000,000 cubic feet of gas per day, handles the vapors from five crude batteries, of total capacity of 90,000 barrels of crude per day. The residue gas from this compression, at 35 pounds per square inch gauge, still containing constituents valuable as motor fuel, is sent to the refinery oil absorption plant for further treatment. The condensate from the compression plant, along with the product from the absorption plant is stabilized by rectification under a pressure of 230 pounds per square inch into a product similar to a good grade of casinghead gasoline. These light products are kept segregated, being handled in specially developed equipment which is described below, and are finally blended back into the naphthas to form finished gasoline.

NAPHTHA STREAMS PURPOSELY RUN HOT

With this system installed, it has been found desirable to run the battery streams at a temperature higher than that to which they could be cooled with the existing condensing equipment. Although this practice increases the evolution of vapors, it has been shown to be economical, because of three very desirable results:

(1) With the greater proportion of the more volatile fractions removed, the naphtha is more stable and can be handled through the subsequent operations of treating and handling with minimum vapor evolution.

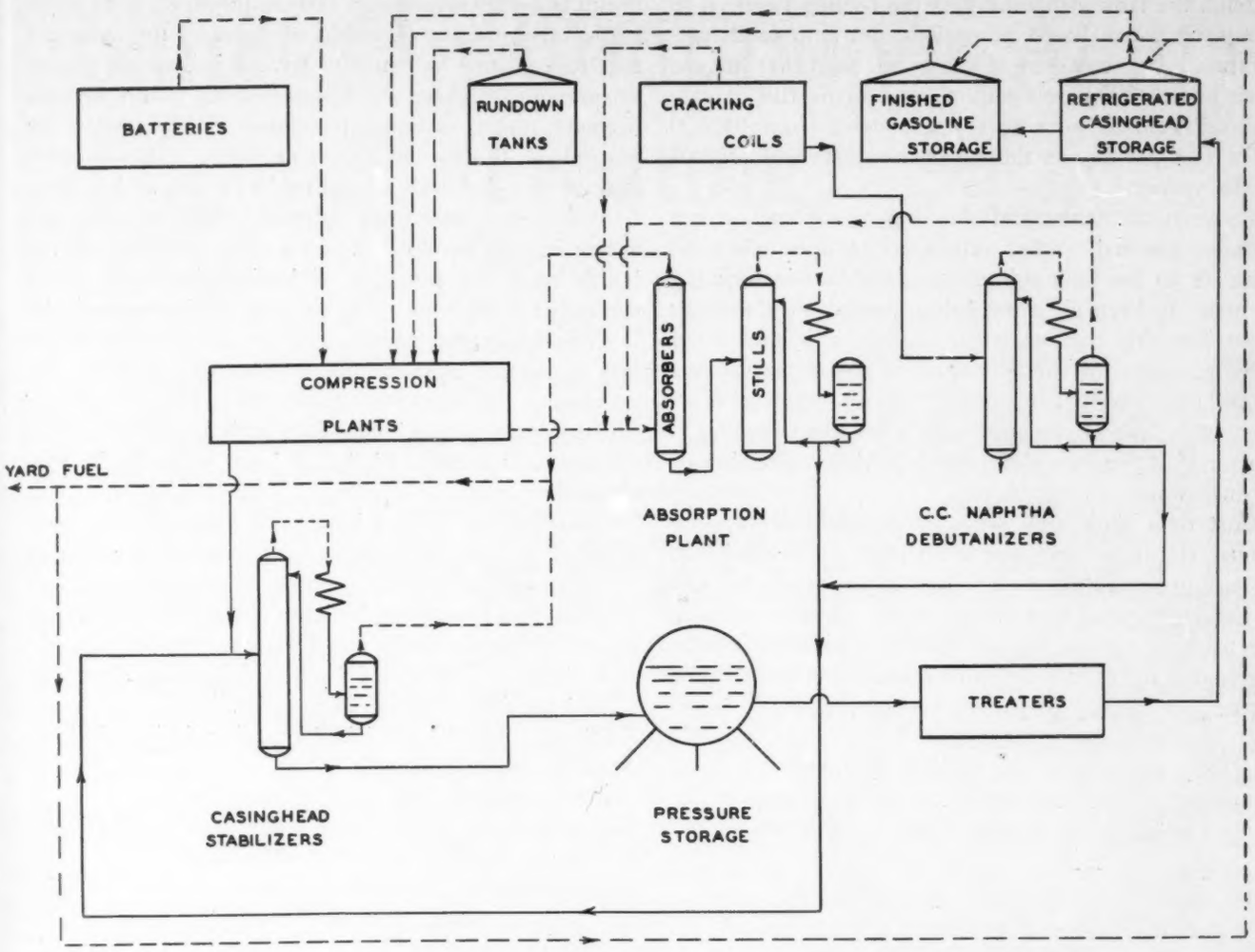
(2) Propane, which is an undesirable constituent in finished motor fuel, because of its excessive volatility, is substantially removed from the naphtha by this treatment. Further, since the recovered light products are subjected to rectification before being put into the final product, this undesirable component is kept out.

(3) As the last few degrees of cooling on the streams are always the most difficult to obtain, the maintenance of higher stream temperatures effects a substantial saving in water consumption.

Of course, for any set of refinery conditions, there is an economical limit to which the stream temperatures should be raised. At Baytown, where the development has been made largely with the use of existing facilities, it has been found economical to increase stream temperatures to about 110°F., at which point the full capacity, with a margin for safety, of the compression system is utilized. The high stream temperatures cause the vapors to the compressors to be very rich. Typical analyses of the residue gas and condensate from this operation are as follows:

	Per Cent by Volume	
	Condensate	Residue Gas
Non-Condensibles	...	48.9
Propane	2.0	9.2
Butanes	27.5	30.9
Pentanes	33.4	9.4
Heavier	37.1	1.6

It should be emphasized that this process is a result of the necessity for utilizing existing distillation equipment. Were new equipment to be built the first still and column of each crude battery would be constructed to operate under a pressure of, say 40 pounds per square inch, and a light cut segregated without necessity for the compression operation. Besides this saving, a further advantage would be the removal of the light fractions under rectifying conditions, allowing a greater



volatility reduction of the naphtha for a given per cent removal than is possible at present.

HANDLING OF CRACKED DISTILLATE

The same principle as was suggested would be applied on a new installation for crude distillation is now in use for stabilizing wild cracked distillates. Here debutanization, as the process has been termed, is effected by rectification under pressure, the non-condensible gas removed going directly to the absorption system, and a highly volatile naphtha overhead cut being combined with the absorption naphtha and compression plant product before stabilization. As much as 15 per cent of the cracked naphtha is thus taken overhead, and the bottoms, having an A.S.T.M. initial boiling point of 125°F. or higher, and a vapor pressure below 200 mm. at atmospheric temperatures can be stored and handled in subsequent operations without evaporation. An additional advantage of this operation, not realized on the straight run naphtha stabilization, is that the light fractions taken overhead by-pass the acid treat given to cracked naphtha. Not only is the expense of this operation saved, but the knock-suppressing tendency of these cracked fractions, which suffers on treating, is conserved.

SEGREGATED STORAGE OF LIGHT FRACTIONS

As indicated on the flow sheet, volatile naphtha fractions, resembling wild casinghead gasolines, are brought to the high pressure rectifying column for stabilization from the following sources:

1. Oil absorption plant.
2. Cracked naphtha debutanizers.
3. Still and tank gas compression plants.

About 18 per cent of the combined streams is taken overhead as a gas, being used for refinery fuel, while the remainder has the characteristics of a good grade of commercial casinghead gasoline, with a gravity of about 90° A. P. I., and an A. S. T. M. recovery of about 92 per cent. This product is run to a working pressure storage consisting of three spherical tanks, of total capacity of 10,000 barrels held at 30 pounds per square inch gauge. From these tanks it is pumped through continuous treaters, being given a caustic wash and doctor sweetening, and is then sent to two 80,000-barrel insulated and refrigerated storage tanks. These tanks, 117 feet in diameter by 42 feet high, are covered with a three-inch thickness of mineral wool insulation and are cooled to and held at about 55° F., the temperature at which the contents have a vapor pressure equal to the atmospheric, by self evaporation to the suction of a compression system. The gas evolution required to do this is that quantity of which the latent heat of evaporation corresponds to the heat conducted

through the tank insulation plus the cooling required to bring any warm liquid pumped in down to tank temperature. By actual test it has been found that the gas from one 80,000-barrel tank three-fourths full of casinghead, held in working storage varies from 400,000 cubic feet per day in the warmest weather to 150,000 in the winter.

Between the tanks and the compressor suction are installed automatic relief valves, set to open when the pressure on the tank side exceeds one-half-inch water. In order to keep air from being drawn into the tanks when pumping out, or when cooling, a second relief valve provides for the admission of gas if the vacuum drops below one-half-inch water. As a precaution, however, each tank is equipped with a breather opening to the air if a pressure differential in either direction of 1.7-inch water is reached.

Cost data show that when large quantities are involved, storage of this type is far more economical than storage in pressure vessels of any type, largely because of the high initial cost of the latter. Yearly costs on the working storage in one 80,000-barrel tank, holding segregated light fractions corresponding in volatility to Grade AA casinghead, at an average gauge of 50,000 barrels, are estimated as follows:

80,000-barrel tank, 117 by 42 feet	$45,000	
Insulation of sides and roof	13,000	
Special lines, valves, instruments	5,000	
Investment	$63,000	
Yearly charges on investment		$9,450
Yearly cost of processing at compression and absorption plants of 300,000 cubic feet of gas per day		8,500
Total yearly charges		$17,950
Cost per year per barrel in storage		35.9c

FINISHED GASOLINE STORAGE

By the initial segregation of the highly volatile and less volatile fractions of naphtha, evaporation is prevented from the latter, and recovered from the former by the methods described. It is obvious that for economic reasons such methods are not applicable to the gross volume of naphtha handled, but only to the small, more volatile, portion. In the final blending operation, however, these segregated fractions must finally be reunited to form finished gasolines.

At Baytown, there are 36 tanks, aggregating about 2,500,000 barrels capacity, in use for the blending and storage of these finished stocks. A system has just been completed and put into operation whereby all vapors expelled from these tanks, due to pumpings and diurnal breathing, are recovered.

The tanks, most of which are of 80,000-barrel capacity, are connected by 12-inch feeders into 15-inch lines, which lead to the suction of a 35-pound compression system. Segregated into convenient groups of six to 10, the tanks in each group are free to equalize in pressure, by vapors flowing from one to another. Between each group and the compression system, however, is an automatic valve opening when the pressure on the upstream side exceeds one half-inch water. As is done on the refrigerated casinghead storage, in case the group of tanks becomes under vacuum, the system is repressured by natural gas through an automatic valve. Likewise each tank is provided with a breather to the air, which functions, however, only when either the intake or outlet gas valves become overloaded, and a dangerous differential builds up. The principle of having the tanks freely connected in as large groups as can be done practically is quite important, as this prevents much evaporation loss by simple equalization, without the necessity for processing the vapors so conserved. The advisability of introducing a gas holder into such a system for this purpose was considered, but did not prove to be economical.

Observations over a long period of time on such tankage, before this system was completed, and also calculations based on average oil movement data and diurnal breathing indicate the great magnitude of losses from volatile gasoline in unprotected tankage. This is illustrated by the following data, obtained over a period of 12 months.

Tankage alloted to finished gasoline, bbls.	2,564,100
Average storage, barrels	1,843,000
Average storage, per cent of capacity	72
Gasoline movements, barrels per year	25,280,000
Barrels moved per barrel storage capacity	9.9
Evaporation loss, barrels per year	105,361
Evaporation loss, per cent per year on average storage	5.72
Evaporation loss, per cent based on gasoline handled	0.86

It is interesting to note that the above losses can be predicted approximately by calculations, as illustrated below.

In addition to the above data, the following assumptions are made:

1. Average vapor pressure of gasoline stored, at storage temperatures is 0.60 atms.

2. Mean annual diurnal temperature variation in vapor space is 35° F.

3. Vapor equivalent of liquid gasoline is 35 cubic feet (Std.) per gallon. From these data can be calculated:

BREATHING LOSS

$$2{,}564{,}000 \times .28 \times \frac{35}{520} \times 365 \times 5.63 = 100{,}000{,}000$$

cubic feet per year Loss due to oil movements

$25{,}280{,}000 \times 5.63 = 142{,}000{,}000$ cu. ft. per year

Total $= 242{,}000{,}000$ cu. ft. per year

The gasoline content is

$$\frac{242{,}000{,}000 \times 0.60}{35 \times 42} = 99{,}000 \text{ barrels per year.}$$

compared with the observed loss of 105,361 barrels. The evaluation of the gross saving if these products

(Continued on page 68)

Notes On Continuous Acid Treating

By EARL PETTY
Bethlehem Steel Company

THE accompanying discussion deals with development work in conjunction with continuous acid treating processes for heavy oils. Upon Mr. Petty's return from Europe, where he spent most of the past year, he was asked by REFINER AND NATURAL GASOLINE MANUFACTURER *to contribute to these columns a discussion which would bring down to date information dealing with this important phase of refining.*

In a communication with the editor Mr. Petty states: "I have never seen a process or an idea so intensively studied and investigated, which at the same time was as highly recommended and supported by each of its investigators, as 'worthy of careful consideration and offers great possibilities for saving and improvement of products.' Each report is closed with enthusiastic recommendations for further work and with tabulated 'pros and cons' comparing the old and the new system.

"With my experience in this country and abroad in the background, and having had access to a number of reports and files which refinery research and equipment manufacturers research and development departments have compiled, I have formed a few ideas and conclusions which may be of interest and value. These I present in the form of notes on continuous acid treating."

DURING the last 10 years there have not been very many exceptions to the assumption that almost every major refinery research organization have made both theoretical and semi-commercial tests of acid treating along the lines of continuous methods using accelerated methods of separation such as various types of centrifugals.

Several companies have compiled files on the subject of continuous treating involving special mixers, centrifugals and filters as the essential equipment of a new treating system. These "hook-ups" have been developed carefully from a laboratory and analytical standpoint and then fitted out in semi-works scales of operation.

Manufacturers of special equipment involved have also spent large sums investigating and developing the fundamental principles of the process with the idea of securing an outlet for their equipment. This later work has been on various scales of development, both in conjunction with and independently of the refiner.

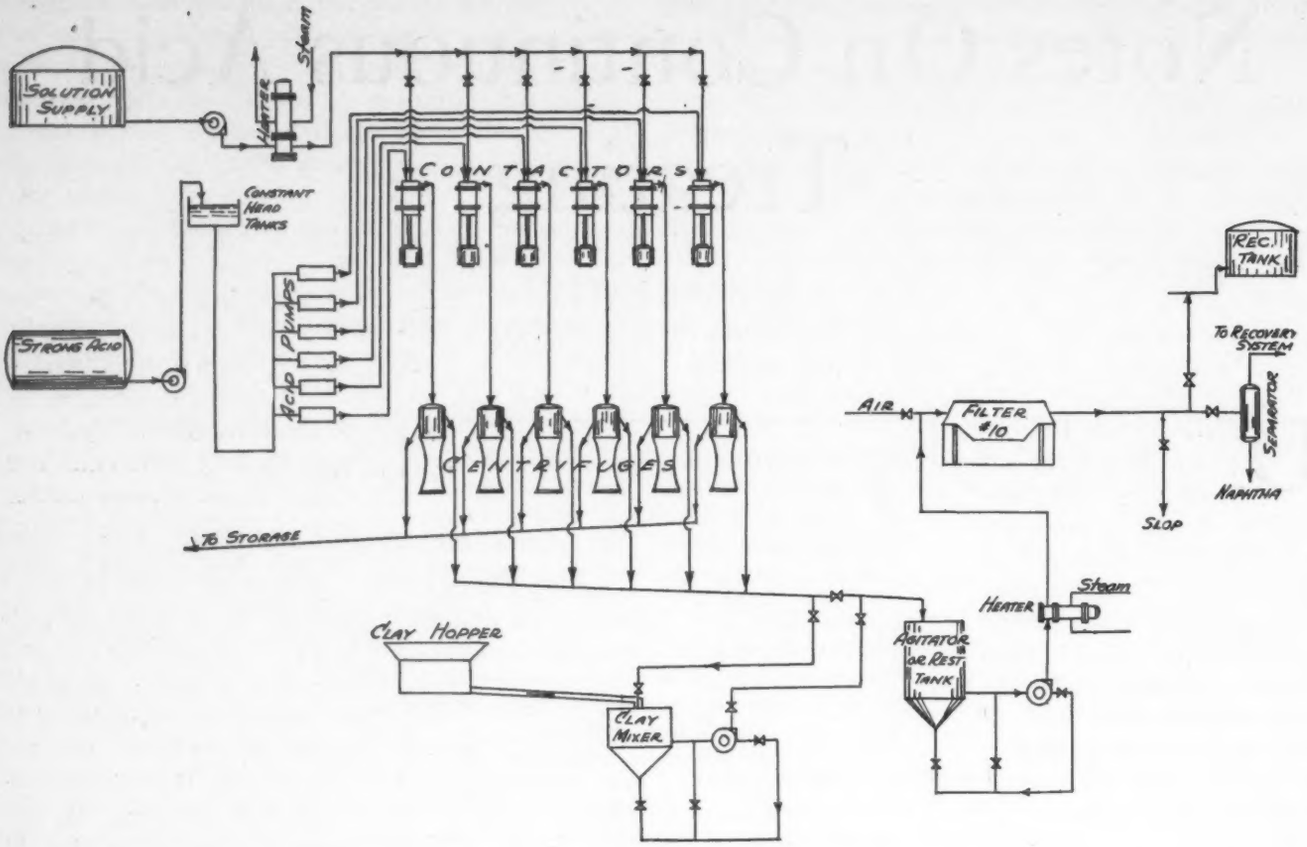
This work has brought to attention many interesting and appealing features and possibilities of the process. To the writer's knowledge, this work began from the large scale centrifugal application standpoint back as far as 1922, when tests were made in conjunction with conventional agitators in trying to reduce the settling time and make a more thorough separation and removal of the sludge and "pepper."

Very often, by prediction, a process may be outlined in a quite radical form. This process, in comparison to the conventional systems, may be very different. The conventional process carries known and recognized handicaps, shortcomings, and bad features. The new process may be founded on true laws of physics and chemistry. Yet in every case there is never an example of a quick change-over. On the other hand, there is always a gradual transition. By the process of evolution the old system is gradually improved. The old process gradually adapts various proven features of the proposed process.

This has been markedly the history of acid treating of the last few years. Gasoline and kerosene treating is carried on almost entirely in closed, and by continuous, methods with mechanical agitation. Separation is left to gravity, but even here special study tended to accelerate methods of separation by extending more surface to a separating plane by placing tanks horizontally instead of vertically, and by making special provisions for feeding and removing the treated liquid from the tanks in order to prevent agitation and cross currents or high velocity stream sections and by use of centrifuges.

In the case of these light products the problem of separation of sludge from the treated products is not so difficult, due to low viscosity of the product being treated and due to the difference in density of the sludge and the treated oil.

When the subject of treating lubricating oils is considered, then factors more complicated enter the picture. Still, the advantages of a continuous system, mechanical



Nine Hundred Barrel Acid Lube Treating Plant

agitation, and complete separation within as short a time as possible, are all very attractive from process and finished product standpoint.

On July 22, 1931, a paper was delivered at the Congress du Graissage, at Strasbourg, France, by Messrs. Backlund & Walch, representatives of the Ste. ALFA-LAVAL, (The Swedish parent company of DeLaval Separator Company). The following statements are extracted from this paper: "The oil industry has been developing for the last few years in all its branches to such an extent that it may justly be claimed that in no other industry have the plants and processes used been so quickly and completely renewed.

"Thus we can mention the case of cracking plants which have been developed to such a large extent, distillation plants which have been improved by the introduction of tubular boilers, fractionating methods involving the use of "bubble towers," and the new dewaxing processes, etc.

"This will give an idea of the great amount of research which has been carried out in the petroleum industry, both in the laboratory and on a manufacturing scale.

"However, one important branch of the industry which we have named the 'refining of petroleum products' has lagged far behind the others. It is not an easy matter to explain the reason for this. At any rate it is clear that the 'refining of petroleum products' is carried out practically the same way as it was towards the end of the last century. The actual methods of treatment have been in use for more than 30 years, and no longer answer the needs of industry, especially where lubricating oils are concerned. This is borne out by the incessant work carried out in this direction."

The problem of removal of coloring material and asphaltic compounds from lubricating oils has not only brought about continuous study and research from a chemical standpoint; but during the last few years has been the main-spring behind the development of vacuum distillation. Vacuum distillation has been the most marked contribution toward the solution of the problem; yet there are compounds of coloring material which have so nearly the same boiling characteristics and range as the oil compounds that it is impossible to commercially make a finished product during the distillation operation.

Several attempts have been made to improve the treating method of lubricating oils by substituting mechanical mixing for the air blowing. Air blowing has been charged with a number of offenses, principally among which are: inefficient mixing; over oxidation; air too dry or too wet; and causing bad odor and bad color due to oxidized compounds re-dissolving in the oil.

Mechanical agitation has made much improvement

but we are still facing the problem of separating the sludge from the oil being refined.

TWO STAGES OF REACTION

Generally we have divided the separation into two stages, viz: sludge separation and "pepper" settling. The latter requires considerable time. It is generally noted that a very good color is obtained at the stage immediately following the first separation and then before the pepper has all separated the color of the refined oil becomes darker. This has been accounted for by the explanation that parts of the coloring materials and compounds have gone back into solution in the oil. This is quite true and has resulted in the general practice of over-treating or treating beyond the color which requires a larger quantity of acid.

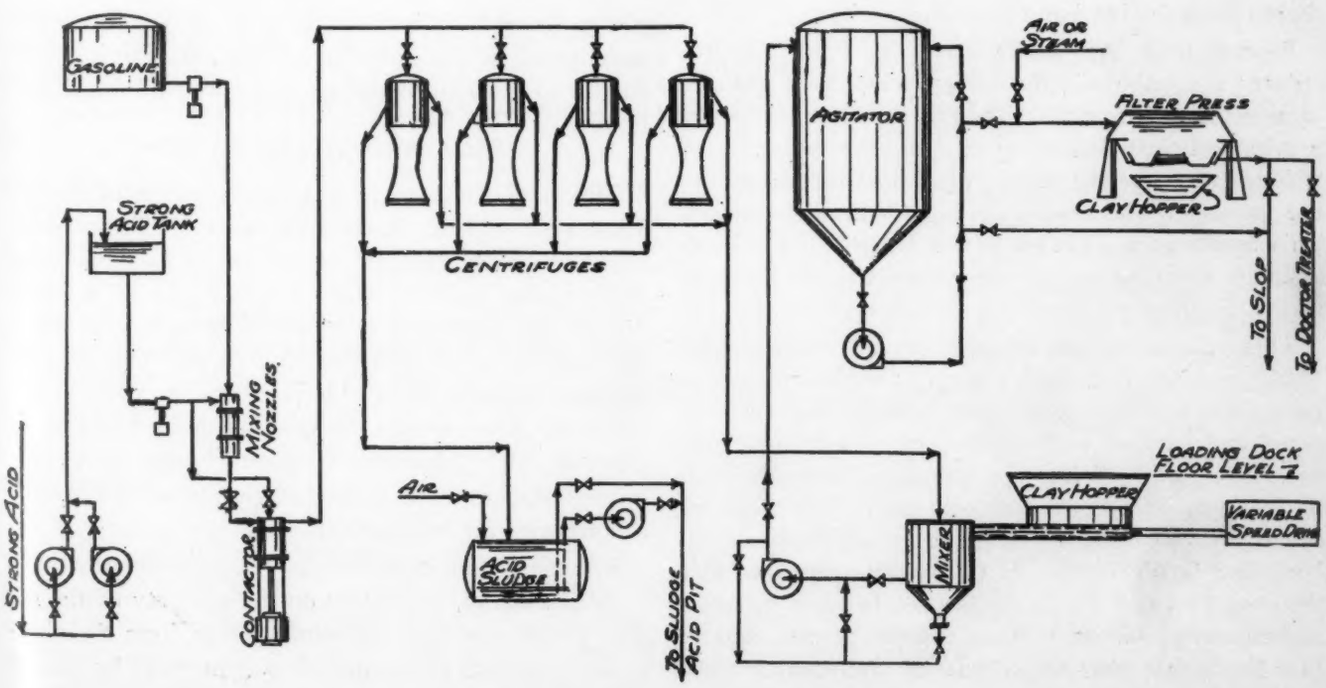
The writer's observation and conclusion has been that there are two stages of the reactions of the acid treatment. These are as follows:

If the correct temperature and time and thoroughness of mixing is used the acid will act purely as a solvent for the asphaltic compounds. If this action could be stopped at this stage and the components separated we would have a refined oil product on the one hand with a trace of acid gas dissolved in it and an acid-asphalt solution on the other hand. The reaction between the dissolved gas and the oil will be very slight as a very small amount of acidity will be carried. On the other hand, there is a very marked reaction ready to take place on the other component which is made up of a large percentage of acid and asphalt. The reaction takes place with considerable temperature rise and with the liberation of tremendous quantities of acid-oil-vapors.

In the regular treating system of batch-agitator operation, the last mentioned reaction is the most offensive feature of this method of treating.

Let us visualize the action that takes place as follows: The air is shut off and there is the sludge thoroughly dispersed throughout the oil in large "pea-like" drops. The agitation is shut off at this stage for best results. The sludge immediately begins to settle and after a few minutes the surface of the oil will become covered with a fine foam and show signs of continual liberation of acid vapors. It is thought that right here is where the pepper begins to form or comes from, viz: the acid vapors coming from the chemical reaction between the acid and the asphaltic materials, passing up through the oil are concentrated enough that another treating action sets up with the pepper sludge as the resulting component. This sludge is much drier and lighter than the first formed sludge. This is due to the fact that it is a product of reaction between SO_2 and the refined oil hence of much lighter density than the first sludge and much more finely divided. Thus we have had two definite reactions take place in the conventional treating process: first, a sludge which is a true solution bond between acid and asphaltic compounds; and, second, the acid-asphalt solution sets up a chemical reaction, SO_2 vapor being one of the resultants and this gas in going up through the treated oil sets up a second reaction which throws a "pepper" sludge.

This conclusion was reached from observations in connection with centrifugal treating where the acid and



Three Thousand Barrel Gasoline Treating Plant

oil were first mixed in a mechanical agitator. The mixture of oil and acid was passed directly into the centrifuge and the sludge separated as the heavier component from the oil as the lighter component. It was noticed that when running at a high temperature, say, 180°F., that a very thin sludge was discharged and a very bright oil. The oil however was very much inclined to froth and foam and upon examination of a quantity of oil taken from a receiver a few minutes after the separation it was found to carry large quantities of fine "pepper." When first discharged from the machine there was no trace of pepper.

Samples were taken and placed between glass slides and carefully observed and after from one to 1½ minutes, minute particles of pepper would begin to form in the bubbles which were first thought to be air bubbles. Further tests were made to determine the relative acidity in oils produced under conditions which gave frothing and forming, and it was found that at high temperatures or with fuming sulphuric acid a considerable quantity of acid gas was dissolved in the oil, and if in high enough concentration, was the cause of the pepper sludge precipitated in the treated oil. Tests were made in which the agitated acid oil and sludge were allowed to stand until the gas component had had time to react with the oil and form the pepper.

It is always noted that this takes more time than the first reaction, since it is a resultant or secondary reaction and must await the liberation of gases by the reaction between the sludge components. This resulted in an oil with a lower acidity; but after the reaction had taken place between the acid and the asphaltic material resulting in the acid vapors, there had taken place a very great change in the character of the sludge: it became drier and harder and much more difficult to discharge from the bowl and from the machine.

Several tests were made after the following procedure: make the mix mechanically and then transfer to a settling tank and withdraw the first sludge by gravity and then decant the oil from the top and centrifuge to remove the pepper which had formed due to the gasses off the sludge passing up through the oil. This results in a slight oil loss in the sludge but gives a sludge when the pepper was accumulated in the bowl that is very dry and hard and difficult to discharge.

This led to a study of temperatures and mixing which would give the least acid vapors. This mixture when passed through the machine gave a sludge that was very liquid and oil free, and an oil carrying very little acid gas. The sludge, from this operation, if studied will be found to be under a decidedly marked chemical reaction, evolving considerable gas and gradually becoming hard and finally brittle. It is thought therefore, that the gases liberated due to the reaction between the acid-asphaltic material solution, causes the pepper, and in turn the high acidity oil, re-solution, and darker color.

These phenomena and result is the greatest justification of the application of the centrifuge to the treating of lubricating oils and has been the most confusing and baffling feature of the problem. In order to facilitate the mechanical operation the separation must be made before the chemical reaction has taken place between the acid and the sludge. The mixing must be done under conditions that will give the minimum tendency to vaporize the acid. This will give a sludge that is easily discharged and an oil that is easily neutralized. It will also give the maximum color with the minimum acid as no over-treating is necessary.

The method of mixing and type of mixers have been subjects of controversy. Too fine a dispersion at too high a temperature will result in total re-solution of the sludge in the oil. Lack of dispersion will result in large globules of acid sludge and inclusion of free acid in the globules. In the first case it will be impossible to make a separation even with centrifuge; and in the second case it will be found that free acid will be separated in the bowl which will result in "acid-seal" which will cut off the discharge of the acid sludge. In either case, too high a temperature or the use of too strong an acid will result in acid gas being dissolved or entrained in the oil, which will later result in a "pepper" or in high acidity and high neutralization requirements.

It is essential to get thorough mixing of acid and oil, but if it is remembered that a process of selective solution is being practiced, in which the acid is the selective solvent and the asphaltic material is to be dissolved, the mixing problem can be easily solved. It is thought that this mixing cannot be by atomizing through spray nozzles or so-called colloidal mixers; neither can it be accomplished by low speed paddles or mixing blades. The most practical mixers are indicated by results given by centrifugal pumps where some slippage is allowed, or with mixers using high speed propellers which discharge through fittings which give fairly sharp angles and turning of the direction of flow, or with impellers similar to centrifugal pumps fitted into mixing chambers.

TIME AND TEMPERATURE

Time and temperature and concentration of the reagent are the next features of control. Temperature enters in mainly from the standpoint of bringing about the optimum viscosity and secondarily as to acceleration of the reaction. It is advisable to use as low a temperature as will give the required viscosity and then increase the time factor to bring about the proper reaction. As noted above, the reaction desired is that of solution only; therefore, the most prominent requirement is proper mixing or the bringing about of thorough contact with the two liquids.

For lubricating oils, the time factor will vary from 6 to 12 minutes, depending on the viscosity of the oil. The viscosity of the oil is referred to from the viewpoint of natural and artificial as influenced by normal temperatures. Temperatures above 140 to 160° F. accelerate the chemical reaction which liberates the vapors

that are dissolved in the oil and produce the objectionable result of high acidity and pepper. Even if the pepper is separated or removed, there is still high acidity if treating is practiced at too high a temperature. With fair consideration to the features outlined above it is possible to make continuous separation giving the following results: a free flowing sludge without oil loss by entrainment; and, a clear, low acidity oil with the use of less acid for a required color.

The neutralization of the treated oil can be carried on by the dry or liquid method. In either method, much less material will be required. By the liquid method less difficulty will be encountered, due to less tendency to form tight emulsions. To follow this method acid treatment with a systematic and conventional contact clay treatment will give the most economical and desirable results.

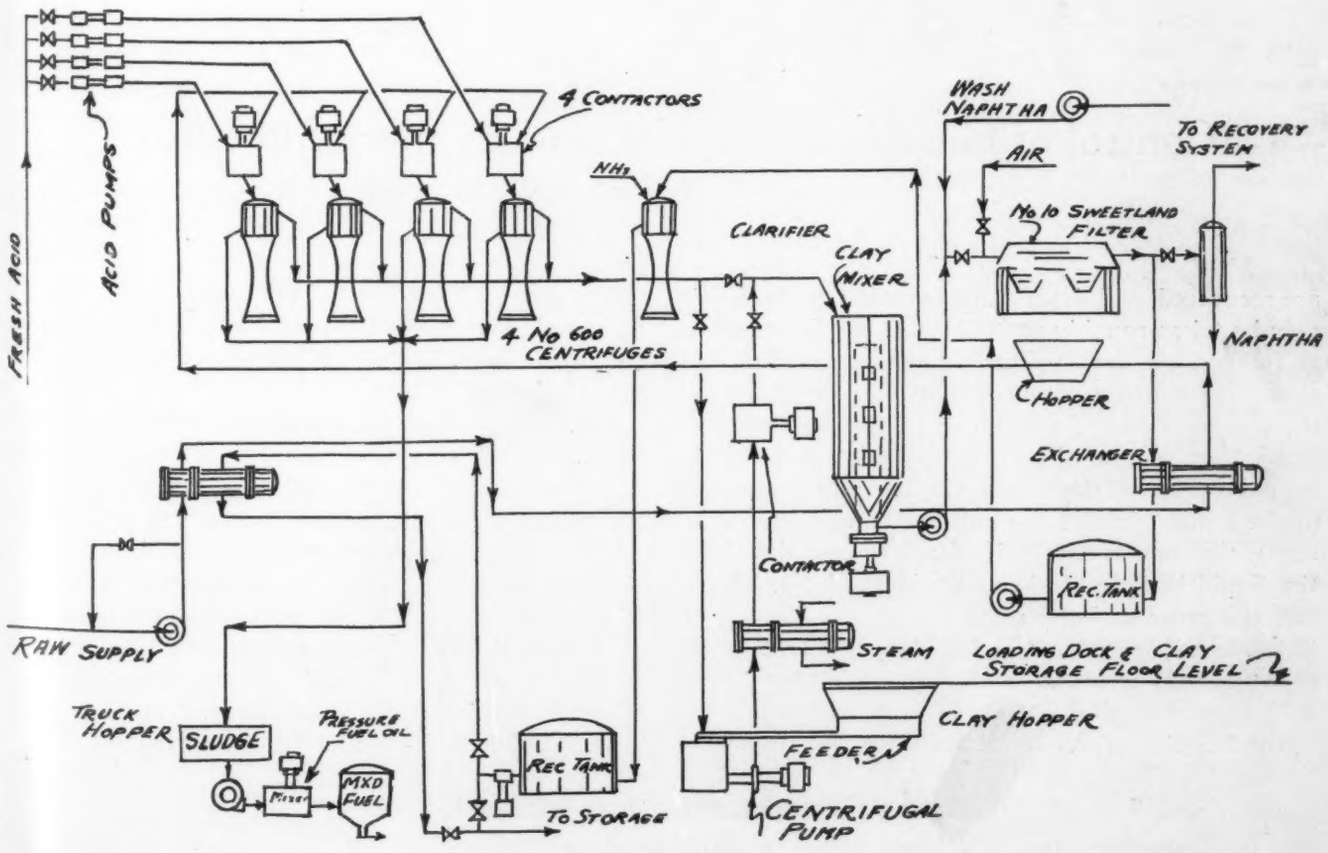
In some cases it may be justified to make the treatment in two stages, but this immediately doubles the mechanical equipment as duplicate mixers, and centrifugals will be required.

It is the writer's opinion that one treat can be carried out that will give the optimum results from acid treating and then the balance of the treatment can be accomplished with the contact clay system. This is justified by the fact that the contact clay equipment will be required even for a small amount of clay for neutralization and a slightly heavier clay treatment can be given with no extra cost except clay. This will give a balance that will be most economical and efficient and give a good finished product.

From a laboratory or semi-commercial plant it is very easy to demonstrate that the same color can be produced with less acid and less loss. However, a very difficult problem has been that of feeding the acid continuously and in unvarying proportions and quantities.

In a batch operation a certain amount of oil is gauged into the agitator and a certain number of pounds of acid added. This gives an exact number of pounds of acid per barrel of oil. From two standpoints the acid must not vary in proportion when treating with the centrifugal continuous method: first, with varying proportions the result will be that of under- or over-treating with an uncertain average or composite sample; second, the solution of acid and asphalt vary in proportion to percentage and with acid going in at a certain rate, say five pounds per barrel, the resulting sludge will be of a certain density, whereas, if the rate is 10 pounds per barrel an entirely different density will result, which will make an entirely different ring size requirement. This is a very serious point from a centrifugal operation standpoint and has caused more trouble, confusion, and false conclusions than any other factor.

Various systems have been tried out, including: constant level tanks and orifice flow feed slots; constant level tanks and proportioning pumps, (liquid driven, and separate electric driven); constant level tanks and



Acid Treating With Clay Filtration

nozzle orifices; and liquid flow controllers, automatic.

Each system has its possibilities but are inherently delicate and are no better than the technique of operation and care. From the extensive laboratory work and research confirming the merits and advantages of the continuous centrifugal process and the rough outline of some of the features of operation, it is readily seen that the process has not been discarded from the standpoint of fundamental soundness. At the same time it is quite evident that the process is highly technical and mechanical and requires the highest precision in machines and operation. It is the latter fact that has been the handicap. The commercial process-hook-up is no better than the machines and the operating technique.

A number of plants have been built that were operable but were discarded due to lack of patience and ingenuity for solving the development of technique and operation. On gasoline a number of successful plants are in operation; but as said above, this is the simplest application of the equipment. The real test lies in the application of the idea to the treating of lubricating oils, and it is in this field that the greatest benefits are to be derived.

Enough theoretical, practical and technical information is available to build operable commercial plants that will not be commercial process mistakes. After the first plants some very great improvements and simplifications will be made from an operating and equipment standpoint, but there will be very few fundamental changes if the design of plant is based on the information at hand in the equipment or refinery company laboratory and development departments. The accompanying flow sheets give some ideas of commercial applications which have been tried out to date.

Earlier in these notes reference was made to a paper presented at Strasbourg. This paper covered the specific results of a commercial plant that has been in operation in the plant of The Standard Nobel, Poland, for over a year, and also from two other commercial plants in Europe which must remain anonymous. A comparative summary of results comparing the method with the old method which was operated under parallel conditions and observations, on a comparison basis using 100 for results obtained by the old method, follows:

	Old method	Contin. Centrifugal
Consumption of sulphuric acid	100	66-72
Loss of oil during the refining treatment	100	70-76
Acid content of centrifuged oil	100	35-50
Loss of oil during the neutralizing stage	100	80-85
Neutralizing material—clay	100	80-85

Control of Gasoline Volatility and Evaporation Losses

(*Continued from page 62*)

are recovered, will depend upon current markets, but may be taken roughly at $162,500. This is arrived at as follows. Analyses of tank gasses show that only about one-fourth is recoverable as liquid products, in the compression and absorption plants together. But as at least half of these gases consist of fractions evaporated from the liquid (as the vapor pressure of the liquid is above one-half the atmospheric pressure), this means that half of the evaporation from the liquid is recoverable as liquid products. Therefore the gross saving is:

50,000 barrels recovered as liquid @ $2.50	$125,000
50,000 barrels recovered as fuel gas @ $0.75	37,500
Total	$162,500

The cost of handling the 242,000,000 cubic feet of gas yearly through both the absorption and compression systems is 7.8c per 1000 cubic feet, or $19,000 per year.

The total cost of installing the necessary equipment on the 2,500,000 barrels of tankage was $290,000. Charges on this at 20 per cent amount to $58,000, giving a total cost of recovery of $77,000 per year, or an annual net saving of $85,000 at which rate the investment would be paid out in 3.5 years. On the above quoted fuel and gasoline prices, this saving is, of course, necessarily affected by changes in the value of gasoline and fuel. In addition to this there is the elimination of hazards due to large volumes of escaping vapor from tankage.

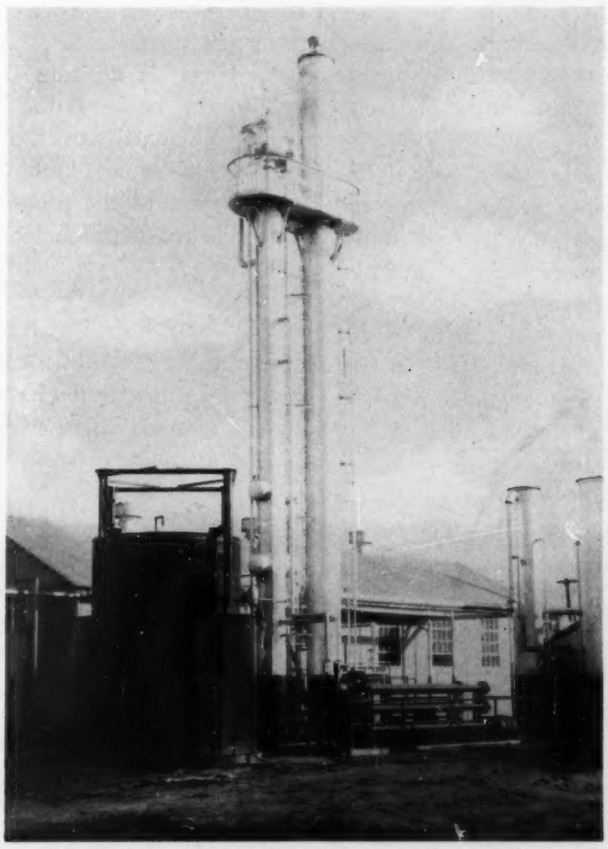
Carbon Dioxide Recovery From Natural Gas

By J. C. ALBRIGHT

UTILIZING that which has been considered a waste product of combustion formed by burning natural gas in the combustion chamber of boilers, Carbonic Products Company, Inc., is successfully manufacturing in quantities both solid and liquefied carbon dioxide in the plant recently completed on North Lewis Avenue, Tulsa, Oklahoma. Many thousands of tons of the raw material from which this new product of natural gas is being manufactured have been going up the stacks of many refineries and natural gasoline plants over the country. Solidified and liquefied carbon dioxide have both been manufactured for several years from various other sources; but it has remained for this company to pioneer the field in producing both forms in commercial quantities from the flue gases when natural gas is used as a fuel for the boilers.

Using a combination of the Girdler and Frick processes, the first in the absorption and purification of the CO_2 vapors, while the second is employed in compressing, liquefying and solidifying the product, this company is making six tons daily. To the casual observer, and more especially to an observer familiar with the routine processes used in the manufacture of natural gasoline, the plant resembles, and is operated in a general way, like a natural gasoline plant

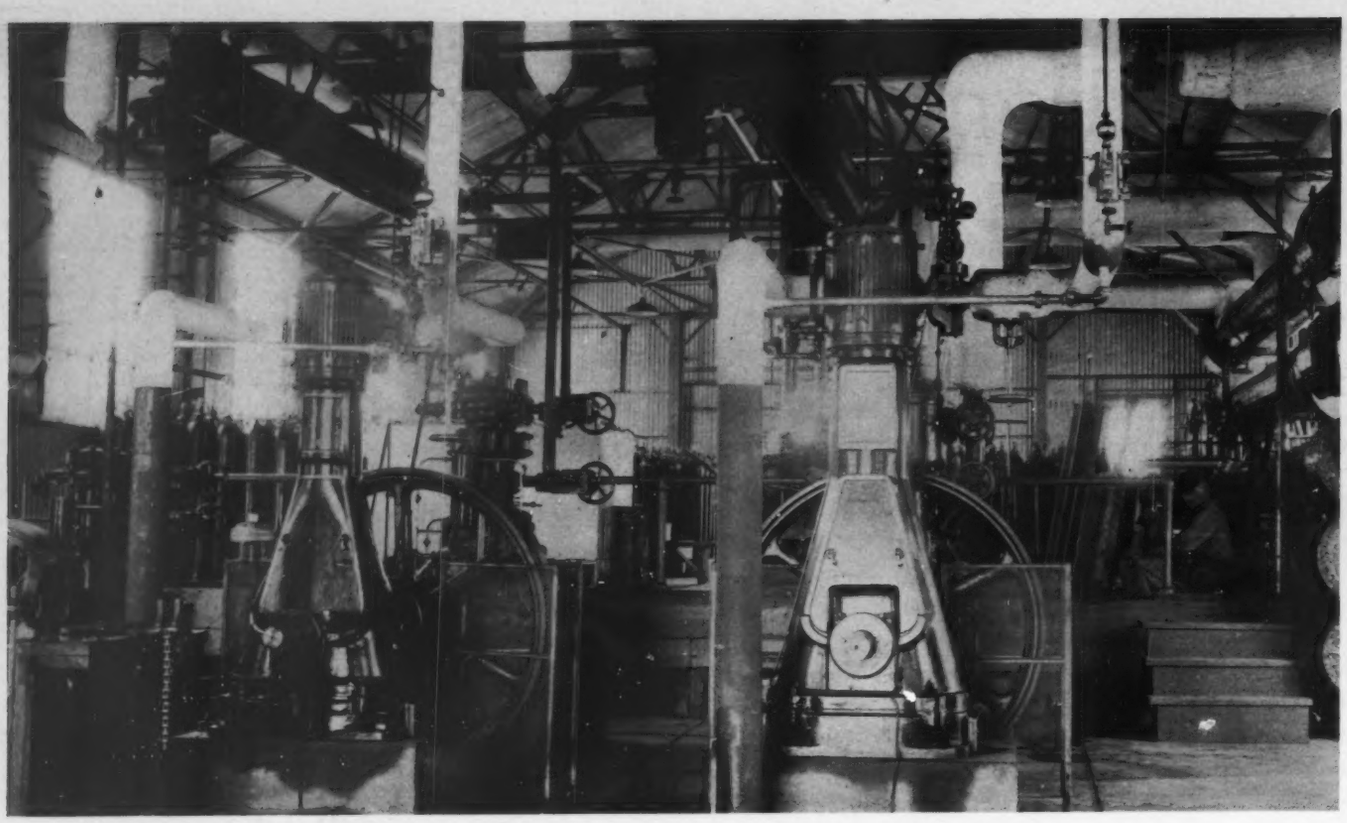
In preparing the gas for combustion, it is brought from one of the mains of Oklahoma Natural Gas Company on the field side of the reducing station through a Neilan boiler fuel regulator. Pressure is reduced to that required for burning by being introduced in the firebox through two Red Star burners manufactured by John Zink Company. A 134 horsepower Hedges, Walsh & Weidner return tubular boiler is used for the dual purpose of securing CO_2 through combustion, and to furnish steam to operate the plant pumps. Steam is used also in the heating coils of the reactivator, or evaporator as it would be called if it were in a gasoline plant.



Absorber and Reactivator in Girdler Process as installed by Jackson Engineering Company for Carbonic Products Company, Tulsa, Oklahoma. Gas holder in foreground to receive CO_2 as it is driven from the rich triethanolamine.

PROCEDURE

Careful mixtures are carried at all times, based upon many analyses of the flue gas, in order that the maximum amount of CO_2 may be produced. An average of 11 per cent may be had from burning the gas available. Other products of combustion are found in this flue gas as generally found in stacks in boiler settings of this character. But to eliminate contamination with air as much as possible, all seams, smoke box doors and the interstices between the bricks of the boiler setting are carefully sealed. Ordinary fire clay and asbestos cement are used, the latter especially around the doors of the smoke box and at other places where iron or steel comes in contact with each other. Fireclay is used to seal the minute holes in the mortar between the bricks. To secure a more efficient and cleaner volume of stack gas, the seams in the smoke stack have been welded from the bonnet



Frick Equipment in Carbonic Product Company's plant at Tulsa

of the boiler to the top of the stack. This precludes the possibility of drawing air in with the flue gases when more suction is placed on this unit. This procedure also prevents rainfall from coming down inside the stack to leak through and forming unsightly rusty looking spots and streaks.

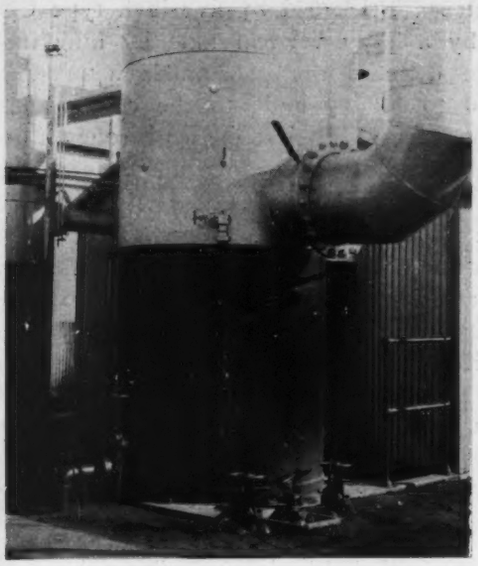
Aldehydes are present in the flue gases, and it has been thought that with the proper setting under the boiler, a temperature high enough may be had in the combustion chamber that these components of the raw material may be eliminated by burning or being converted to other gases that will not be harmful or expensive to remove by scrubbing. Scrubbing the CO_2 is necessary at the present time to remove those gases in order that a pure product may be formed into solid or liquid form which will meet with trade specifications.

The boiler is equipped with draft regulators, manual in type, operated by the boiler fireman when by observing the draft gauge he finds manipulation necessary. Careful stack draft is maintained by balancing the draft dampers in the stack itself and the damper or butterfly valve in the duct leading from it to the plant. Water is fed to the interior of the shell of the boiler through the utilization of a Dean Brothers four by five by seven inch simplex steam pump governed by a Copes boiler feed water controller placed on the supports of the boiler setting. Preheating of the water is accomplished through the use of a Tulsa Type preheater and exhaust steam condenser, saving practically all the steam from the pumps and heating the water to a temperature sufficiently high that fluctuation of fuel feed is eliminated. Average constant temperatures are maintained as nearly as possible so that the rate of combustion is not altered. If the rate of feed through admitting more gas to the burners due to feeding cold water in the boiler fluctuated over relatively wide amounts, the equilibrium is lost and careful manipulation is again necessary until the proper temperature is attained that will give the maximum amount of CO_2 in the flue gas.

ARRANGEMENT

A 24-inch steel duct leads from the boiler stack where it is retained to that member just above the bonnet by a riveted connection which was afterwards welded. This duct leads to the first water scrubber, two of which are placed near the compressor and pump room of the absorption plant. A Roots, Connersville Wilbraham rotary blower, 12 by 18, type SD is utilized for this unit, driven by a Troy seven by seven inch SHT vertical steam engine. These two units are directly connected and as the steam engine is governor operated, the speed is constant after the combined units are placed in operation.

In the compressor room, located in a Braden all steel building, are the water circulating pumps, the Leeds & Northrup CO_2 recorder and a Neilan temperature recorder controller. Three Dean-Hill 2½ inch single stage centrifugal pumps are set on the foundations, so connected that the third is a stand-by unit which may be used on either of the two water circuits. They are driven by Lewis Allis direct connected type X self start-



All gas from boiler stack enters through large steel transmission duct leading to this water scrubber. Removal of ash, soot and water soluble impurities.

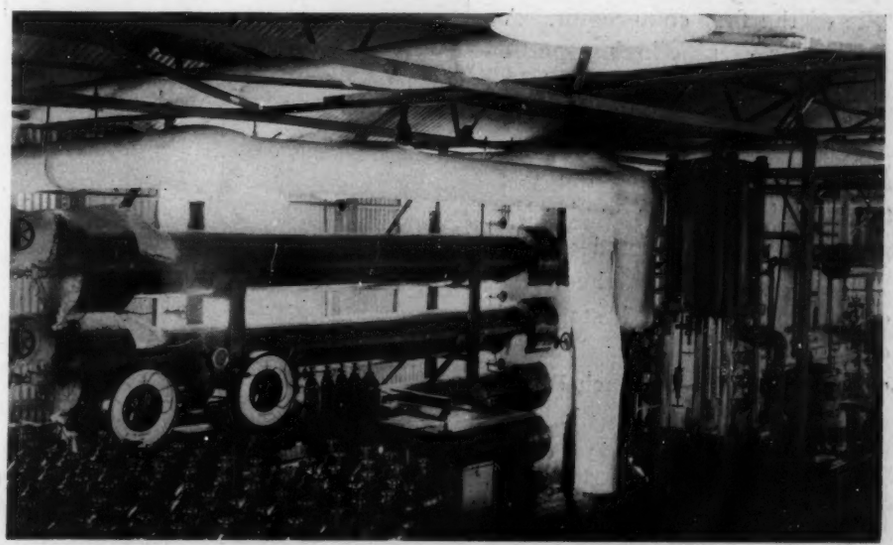
ing motors controlled by Cutler-Hammer switches, Designed to operate at 1750 revolutions per minute, and at this speed, the pumps are capable of handling 130 gallons of water per minute. One of the pumps is used in circulating the water from the stack gas scrubbers, taking suction on the louver tower pit and discharging it near the top of both scrubbers where it falls in counter current to the ascending flue gas. This water is circulated continuously and not permitted to waste. The other pump takes suction in the same manner and from the same common source, discharging over the reflux condenser of the reactivator in exactly the same manner that it is circulated through these units on a still or dephlegmator in a gasoline plant. Part of the stream of water from this pump is diverted to the intercoolers and condensers in the CO_2 compressor room.

The flue gases receive the initial scrubbing in the first vertical scrubber by having been drawn from the stack by the suction of the rotary blower. The discharge line leads from this unit to the second scrubber, the gases passing through to the absorber. Entering this unit near the base, the gas comes in contact with the stripped triethanolamine. The absorber, 65 feet in height having 36 plates was originally only 50 feet tall, but to give a longer time contact between the descending fluid and the flue gas containing the desired CO_2, 15 additional feet were added by placing an extension on the top of the original tower. As the gas enters the base of the absorber in the same manner as that in ordinary gasoline practice, it flows upwards to leave at the top, the stripped gas being vented to the atmosphere.

TRIETHANOLAMINE DILUTED

The triethanolamine placed in the system is diluted to approximately 50/50 with water, that ratio between the absorbing fluid and water having been found to give the desired results. However, a lower ratio of the amine to water performs in a satisfactory manner until the amount stands below 30/70, when an excessive circulation must be maintained to strip the CO_2 from the stack gas. Circulation is maintained in this manner. This lean or stripped liquor is released from the lower compartment of the reactivator through a Neilan liquid level control to pass through a series of heat exchangers and coolers. These exchangers and enclosed tubular coolers as well as the other units of the absorption system were manufactured by Jackson Engineering Company and are made of iron and steel. No brass or copper alloys are permitted to be placed in any point of the amine circuit. After passing through these exchangers and coolers, similar in construction and identical in operation to those placed in gasoline plants, the fluid or stripped liquor enters the suction of a Dean Brothers four by five by seven inch simplex steam pump, the speed of which is controlled by a Mason pump governor. The speed of this pump has been synchronized with the flow of liquid from the reactivator so that it may not create undue suction on the lines leading from the units in the evaporating or reactivating equipment in the plant.

Three of these pumps, all of the same dimensions, are placed near the base of the absorber and reactivator, one of which is always in readiness to be utilized as a standby unit on either the rich or stripped liquor circuits. The pump handling the stripped liquor from the heat exchangers and tubular coolers discharge it to a point near the top of the absorber column. It has been reduced to a temperature having been found to give better absorbing than if the fluid were cooler. Flowing downward over the numerous plates in the



Snow machine and equipment to place liquid CO_2 in cylinders in plant of Carbonic Products Company in Tulsa. Snow machine with hydraulic press. Scales to determine amount of liquid in cylinders, and to check uniformity of blocks of solid CO_2.

absorber, the amine removes the CO_2 from the flue gas entering the base. It is removed through the action of Neilan liquid level controller to the rich liquor pump. This pump discharges the fluid back through the exchangers to enter the top of the reactivator.

As this rich liquor carrying the CO_2 in solution, enters the upper section of the reactivator in the same manner as the rich oil in a gasoline plant enters the evaporator, it travels downward refluxing the ascending vapors which consist of water vapor and CO_2 gas. As it descends, it becomes heated to higher temperatures until it comes in contact with the first bundle of heating coils. Two of these elements are placed in the reactivator, one above the other and they may either be operated in series in so far as steam is concerned, or they may be operated singly or in parallel. This depends entirely upon the amount of steam available and the degree of heat necessary to liberate the CO_2 from the rich liquor coming in contact with them.

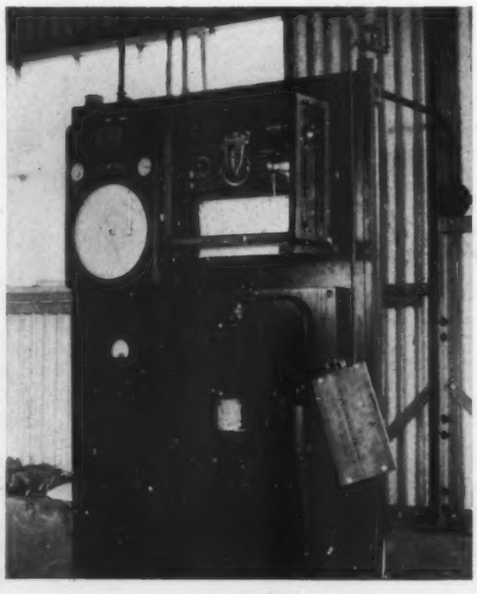
ASCEND IN COUNTER CURRENT

The vapors rising consisting of water and CO_2 always ascend in counter current to the descending fluid until they reach the top of the reactivator. They then enter a tubular enclosed reflux condenser, water cooled to reduce the water vapor to liquid. This falls back down into the stream of descending rich liquor to maintain an even balance between water and triethanolamine. The liberated CO_2 gas leaves the top of the reflux condenser through a line that has been placed inside the shell of the reactivator, emerging near the base. It extends under the surface of the ground to the lower part of the gas holder through liquid traps. This gas holder or gasometer forms a reserve storage for the CO_2 in order that there may be no vacuum on the reactivator.

A motor driven Roots Connersville Wilbraham size 289 rotary blower removes the CO_2 from the gas holder passing it through the purifier and deodorizer. Leaving this unit as 99.9 per cent pure CO_2 gas, it is admitted to the intake of the first compressor. A Frick nine by nine inch, type G, vertical twin driven by an eight by 10 inch Troy vertical single cylinder steam engine directly connected, compresses the gas to approximately 80 pounds. No condensation is effected as the gas passes through the inter cooler to the second stage of compression. Another Frick machine, 4½ by five inch direct connected to a Troy nine by nine-inch vertical engine compresses the gas to approximately 400 pounds, delivering it through the second set of intercoolers to the third and final stage of compression. This last machine is also a Frick 3¾ by five inch type CO_2 also driven by a Troy vertical nine by nine inch steam engine delivers the gas under a pressure of from 1100 to 1200 pounds to the conventional horizontal condensers. It first passes through three vertical scrubber filters to remove all traces of lubricating oil which may have been carried on with the gas from the compressor cylinders. As the critical temperature of CO_2 is approximately 88.5°F., the cooling water from the louver must always be low to effect condensation of the gas in the condenser sections.

The liquefied carbon dioxide is passed through the lower section of the condenser to be received in another section below before passing to the liquid coolers and heat exchangers. As the liquid enters these sections, three of which are liquid coolers while the other two are heat exchangers a part of the liquefied CO_2 is expanded through Crane needle valves to the shells of the section to maintain a low temperature. This expanded gas is passed back to the intake of the three machines, the greater part of which enters the suction line of the first compressor. The five sections of final coolers and heat exchangers are heavily insulated with cork furnished and applied by Armstrong Cork Company, while the pipe lines leading from them are insulated by L. Mundet Sons Company. The expanded gas is recompressed and returned.

The finally cooled and refrigerated liquid is now in condition to be manufactured into snow, or placed as liquid in the cylinders for shipment. If solidification is desired, the purified and refrigerated CO_2 is admitted to the regulation Frick snow machine. This machine prepares the snow in solid form in 10-inch cubes by compressing after being formed in the snow chamber. Two rams, both hydraulic operated, one below compressing while the upper seals and maintains a barrier against the release of the snow in the chamber.



Leeds & Northrup continuous CO_2 recorder showing at all times the amount of CO_2 gas in duct from boiler stack

Solidified CO_2 formed in the 10 by 10 by 10 inch cubes, weigh approximately 53 pounds. Fifty pounds is the standard weight, but to provide for shrinkage, this company has added the extra amount in order that the ice will reach destination full weight. Storage boxes lined with thick layers of cork are built into the room in which the ice is placed as manufactured. When making solidified, a cube comes from the press each six minutes. If the material manufactured by the company is to be placed in cylinders, they are placed on the loading rack set on a pair of Toledo scales, the reading of which as the liquid enters them, enables the operators to fill to standard capacity.

Effect of Concentration on Adsorption

HAROLD J. TORMEY, FRED L. KOCH, JAMES J. O'CONNOR
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IT has long been known that the concentration of a substance to be adsorbed from its solution in water has a marked influence on the extent of the adsorption. For most substances, adsorption has been found to be highest at low concentrations.

In view of the fact that practices in percolation and contact filtration vary markedly in different refineries, it seemed desirable: (1) to attempt to determine the affect of concentration of stock in various solutions and to ascertain for each solvent studied the concentration at which the greatest amount of adsorption took place; and (2) to attempt to determine the adsorbing efficiency of various substances whose surface properties as well as economic availability indicate them to be possible substitutes, either in whole or in part, for the comparatively expensive clays now widely used for this purpose.

Accordingly, the percentage of color removal produced by percolation through a standard clay in a laboratory filter of solutions of a lubricating stock of various concentrations in both benzene and naphtha has been determined. Using the lubricating stock in the concentrations in benzene and naphtha found most efficient in the above described study, the comparative efficiencies of various absorbent mixtures, expressed in terms of percentage color-removal, have been determined.

In general, the methods for measuring the amount of adsorption fall into two classes: (1) methods involving the measurement of change in concentration of the substance removed from the solution by adsorption; and, (2) methods involving the measurement of the extent of the reaction as reflected in energy changes. Measurement of change in color, viscosity, etc., are examples of methods of the first class while measurement of heat of wetting is an example of a method of the second class.

Since methods of the first class are more widely used and are, perhaps, better known to most refinery chemists and since the measurements required by these methods are more readily made, a method of this class, namely, change in color, was selected for the determination of the amount of adsorption in these studies.

The large variety and great complexity of substances removed in part from lubricating stock by adsorption make difficult a direct quantitative determination of these substances before and after adsorption; consequently, the measurement of the concentration of these substances before and after adsorption has taken place is made by measuring some property which depends on these concentrations and which depends on the joint presence of all the substances which it is desired to remove. Such properties are color, viscosity (to a limited extent) etc.

Solutions of a lubricating stock in the solvents and concentrations shown in the table were filtered through 65 grams of a standard clay in the small lab-

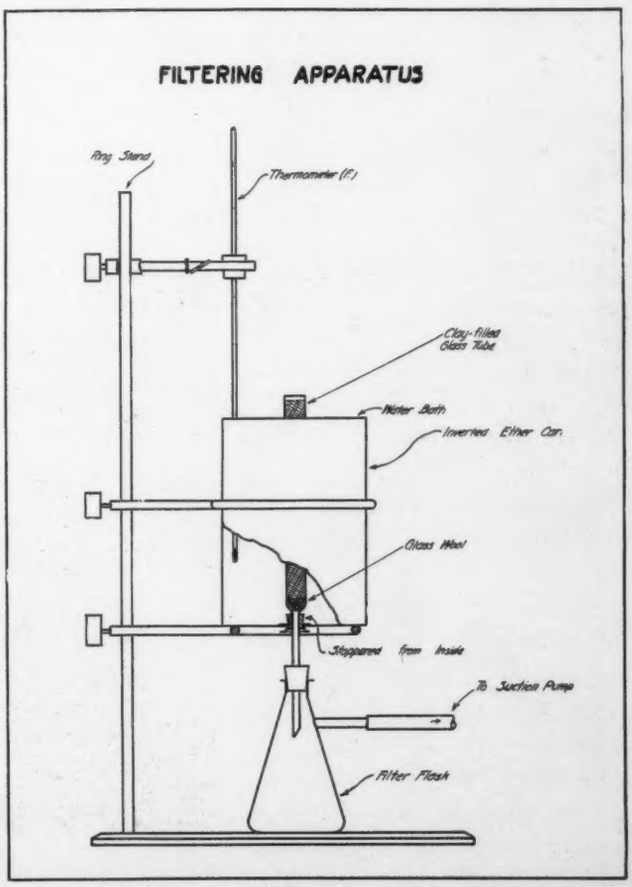


FIGURE 1

oratory filter illustrated in Figure I. The clay used was 40-60 mesh, this being the mesh commonly used for percolation filtration. The solution under test, preheated to 100° F., was slowly poured onto the clay and allowed to stand until a penetration of about two inches was obtained. The tube holding the clay was surrounded by a water bath maintained throughout at 100° F. Pressure was then slowly applied, and after ascertaining that channeling was not taking place, the pressure was increased to a moderate value. The time noted in the table is the time required for 100 c.c. of the filtrate to collect in the flask.

The iodimetric method as described in Bulletin Number 25 of the Kansas City Testing Laboratory was employed to determine the color before and after filtration. The procedure is as follows: One cc. of the solution under test is placed in a color tube and diluted to 100 cc. with the same solvent with which the solution had been made (i.e. naphtha or benzene). In the second of the two color tubes was placed one cc. of the standard iodine solution. Enough distilled water was added to the standard iodine solution in the color tube to bring its color to a match with that of the oil solution. The percentage of color removal was then calculated according to the following equations:

(1) $\text{Color} = l\,(d+1)$

l = number of milligrams of iodine in 100 cc. of water in the tube containing the diluted iodine.
d = # cc. benzene to 1 cc. of oil.

(2) $\%\ \text{Removal} = \dfrac{\text{Iodimetric color original} - \text{Iod. color bleached}}{\text{Iodimetric color original X 100}}$

The second phase of this investigation consisted of a comparison of the efficiencies of various adsorbents of the lubricating stock in benzene and naphtha in the ratio found most efficient in the previous study. For benzene the ratio was 90% of stock to 10% of benzene. For naphtha, the ratio was 80% of stock to 20% of naphtha. The percentage of color removal was taken as an index of adsorption efficiency.

The results obtained in the study of the affect of concentration on adsorption are expressed in terms of percentage color removal in Table I. Greatest adsorption was found to take place in the more concentrated solutions; the greatest percentage of color removal occurring in the solution containing 90% of stock to 10% benzol and in the solution containing 80% of stock and 20% of naphtha. The curve obtained by plotting percentage removal against percentage of stock in solution for both the benzene and naphtha solutions shows a pronounced trend toward greater adsorption in the more concentrated solu-

Table No. 1

Effect of Concentration on Adsorption in Benzol Solutions of Lubricating Stock

Description Clay	Adsorbent Time (In Minutes)	Size	Weight (In Gm.)	Solvent Benzol Per cent	Stock Per cent	Test Method	Iodimetric Color Before	Iodimetric Color After	Per cent Removal
Standard	50	40-60	65	95	5	Filtration	5537.23	3678.16	33.6
Standard	70	40-60	65	90	10	Filtration	5674.32	3636.36	35.8
Standard	90	40-60	65	85	15	Filtration	5411.81	4666.27	13.7
Standard	90	40-60	65	80	20	Filtration	4705.57	4056.87	13.8
Standard	110	40-60	65	75	25	Filtration	3950.10	3471.30	12.1
Standard	85	40-60	65	70	30	Filtration	3645.66	2864.68	20.7
Standard	90	40-60	65	65	35	Filtration	3384.04	2855.64	15.6
Standard	98	40-60	65	60	40	Filtration	4148.34	2557.23	38.4
Standard	67	40-60	65	55	45	Filtration	4696.50	2703.30	42.5
Standard	205	40-60	65	50	50	Filtration	4855.60	3148.18	35.2
Standard	75	40-60	65	45	55	Filtration	6388.37	3246.97	49.3
Standard	90	40-60	65	40	60	Filtration	5915.71	3576.59	39.5
Standard	65	40-60	65	30	70	Filtration	5885.77	3772.05	35.9
Standard	105	40-60	65	20	80	Filtration	5903.64	2130.32	63.9
Standard	95	40-60	65	10	90	Filtration	5825.92	1977.35	66.0

Effect of Concentration on Adsorption in Naptha Solutions of Lubricating Stock

Description Clay	Adsorbent Time (In Minutes)	Size	Weight (In Gm.)	Solvent Naptha Per cent	Stock Per cent	Test Method	Iodimetric Color Before	Iodimetric Color After	Per cent Removal
Standard	65	40-60	65	95	5	Filtration	6656.67	2138.93	68.0
Standard	75	40-60	65	90	10	Filtration	4435.56	1238.76	72.2
Satndard	65	40-60	65	85	15	Filtration	4206.97	918.61	78.0
Standard	105	40-60	65	80	20	Filtration	3792.40	713.57	81.1
Standard	70	40-60	65	75	25	Filtration	3407.46	582.54	82.9
Standard	65	40-60	65	70	30	Filtration	4041.10	548.34	86.7
Standard	57	40-60	65	65	35	Filtration	4313.35	579.42	86.5
Standard	100	40-60	65	60	40	Filtration	4788.27	891.50	81.4
Standard	70	40-60	65	55	45	Filtration	5530.50	1087.34	88.5
Standard	90	40-60	65	50	50	Filtration	5253.60	1345.24	74.0
Standard	75	40-60	65	45	55	Filtration	5511.08	1068.58	80.5
Standard	65	40-60	65	40	60	Filtration	7668.40	949.32	87.7
Standard	65	40-60	65	30	70	Filtration	6776.10	834.07	82.7
Standard	105	40-60	65	20	80	Filtration	7750.00	703.32	90.7
Standard	95	40-60	65	10	90	Filtration	6924.81	698.09	90.0

Table No. II

Relative Efficiencies of Different Adsorbents on Benzol Solutions of Lubricating Stock

Description	Adsorbent Time (In Minutes)	Size	Weight (In Gm.)	Solvent Benzol per cent	Stock per cent	Test method	Iodimetric Color Before	Iodimetric Color After	per cent removal
50% C-50% I	100	40-60	65	10	90	Filtr.	5825.92	2351.94	40.40
50% C-50% S	105	40-60	65	10	90	Filtr.	5825.92	2572.16	44.20
50% C-50% Sa	95	40-60	65	10	90	Filtr.	5825.92	2560.05	44.00
50% C-50% A	90	40-60	65	10	90	Filtr.	5825.92	3597.29	61.70
50% I-50% S	125	40-60	65	10	90	Filtr.	5825.92	3431.79	58.80
50% I-50% Sa	180	40-60	65	10	90	Filtr.	5825.92	4765.56	81.90
50% I-50% A	120	40-60	65	10	90	Filtr.	5825.92	3351.54	57.50
50% S-50% Sa	170	40-60	65	10	90	Filtr.	5825.92	4587.28	78.70
50% S-50% A	190	40-60	65	10	90	Filtr.	5825.92	5242.33	89.80
50% A-50% Sa	185	40-60	65	10	90	Filtr.	5825.92	4915.31	84.40
25% C-75% I	175	40-60	65	10	90	Filtr.	5825.92	3770.16	64.80
25% C-75% S	128	40-60	65	10	90	Filtr.	5825.92	4033.32	69.20
25% C-75% Sa	106	40-60	65	10	90	Filtr.	5825.92	4063.05	69.80
25% C-75% A	115	40-60	65	10	90	Filtr.	5825.92	4169.86	71.50
75% C-25% I	127	40-60	65	10	90	Filtr.	5825.92	2100.89	36.10
75% C-25% Sa	160	40-60	65	10	90	Filtr.	5825.92	2129.52	36.80
75% C-25% A	155	40-60	65	10	90	Filtr.	5825.92	2383.78	41.00
75% C-25% S	173	40-60	65	10	90	Filtr.	5825.92	2731.82	46.80

ADSORBENTS: C—Clay (Standard). I—Industrial Waste Product. S—Sand. Sa—Sawdust. A—Asbestos.
METHOD: Filtr.—Filtration.

tions; although these curves show numerous irregularities it is believed that these may be eliminated by further refinements in the experimental procedures, particularly with respect to time of contact with the adsorbing agent.

The results obtained in the determination of the relative efficiencies of various adsorbents on the solutions of the stock in benzene and naptha at the concentrations found most efficient in the above study are expressed in terms of percentage color removal in Tables II and III. With reference to the benzene solution of the stock (90% benzene and 10% stock) it will be noted that seven of the mixed adsorbents gave greater percentage color removal than the standard clay. Of these seven admixtures, particular interest attaches to that containing 50% of sand and 50% of an industrial waste. Percentage of color removal for this adsorbent mixture was found to be 81.9% as against 66% for the standard clay. Although this particular admixture was not the most efficient of those tried (see table) nevertheless it seems worthy of further study because of the availability of both of the materials in large quantities and at low costs. With reference to the naptha solutions of the stock (80% stock and 20% naptha) it will be noted that none of the mixed adsorbents can favorably compare with the standard clay.

The industrial waste referred to above is the waste material resulting from the manufacture of alkali by the ammonia soda process.

Table III.

Relative Efficiencies of Different Adsorbents on Napththa Solutions of Lubricating Stock

Description	Adsorbent Time Minutes	Size	Wt. (in Minutes)	Solvent Naphtha per cent	Stock per cent	Test method	Iodimetric Color Before	Iodimetric Color After	per cent removal
50% C-50% I	125	40-60	65	20	80	Filtr.	7750	988.28	12.76
50% C-50% S	140	40-60	65	20	80	Filtr.	7750	1046.56	13.49
50% C-50% Sa	135	40-60	65	20	80	Filtr.	7750	1057.72	13.62
50% C-50% A	130	40-60	65	20	80	Filtr.	7750	1342.92	17.68
50% I-50% S	133	40-60	65	20	80	Filtr.	7750	2880.52	37.10
50% I-50%Sa	151	40-60	65	20	80	Filtr.	7750	3850.20	49.70
50% I-50% A	168	40-60	65	20	80	Filtr.	7750	3211.60	41.50
50% S-50% Sa	174	40-60	65	20	80	Filtr.	7750	3542.68	45.07
50% S-50% A	160	40-60	65	20	80	Filtr.	7750	4438.04	57.20
50% A-50% Sa	147	40-60	65	20	80	Filtr.	7750	3974.20	51.30
25% C-75% I	126	40-60	65	20	80	Filtr.	7750	1424.76	18.36
25% C-75% S	162	40-60	65	20	80	Filtr.	7750	1452.04	18.73
25% C-75% Sa	154	40-60	65	20	80	Filtr.	7750	1678.96	20.80
25% C-75% A	171	40-60	65	20	80	Filtr.	7750	1800.92	23.20
75% C-25% I	132	40-60	65	20	80	Filtr.	7750	885.36	11.43
75% C-25% Sa	163	40-60	65	20	80	Filtr.	7750	863.04	11.13
75% C-25% A	145	40-60	65	20	80	Filtr.	7750	967.20	11.20
75% C-25% S	137	40-60	65	20	80	Filtr.	7750	1140.80	14.72

ADSORBENTS: C—Clay (Standard). I—Industrial Waste Product. S—Sand. Sa Sawdust.
METHOD: Filtr.—Filtration.

Osterstrom Vapor Phase Treating System

By DR. W. W. HOLLAND

RAPID developments in the automotive industry have created many problems in the preparation of correct motor fuels. Perforce, this became the responsibility of the petroleum industry. Engines of the earlier type operated at low speeds and low compression ratios, and the so-called "straight run" gasolines were satisfactory as fuels. These engines, however, were very inefficient. As motors were improved, the tendency was toward lighter construction, higher speed and higher compression ratios. Immediately difficulties were encountered with the grades of available fuel, due to their inclination to "knock." This resulted in reduced power and was also detrimental to the motor itself.

At this point the petroleum industry accepted the problem of developing a satisfactory motor fuel; and the means of accomplishing the desired results are well known. Among the early attempts to produce a suitable fuel was the cracking of heavy hydrocarbons in the vapor phase, as it was known that this type of fuel was very efficient because of its B.t.u. value and its inherently high anti-knock properties. However, the many difficulties encountered caused this process to be abandoned temporarily, as it could not compete economically with the liquid phaseprocess which was perfected about the same time.

In 1926, Gyro Process Company, anticipating the demands of the future, undertook to solve the problems of vapor phase cracking, the most important of which was perfecting a satisfactory method of treating the cracked distillate whereby a water-white stable product of high anti-knock value could be obtained.

Distillate of this character contains 40 to 45 per cent of aromatics, 45 to 50 per cent of olefines and the balance naphthenes, with paraffines almost entirely absent. It was apparent that distillate of this nature would not lend itself readily to the conventional methods of treating and refining without sacrificing some of its desirable qualities.

Treatment with sulphuric acid affected the anti-knock properties and resulted in excessive treating losses.

Treatment in the vapor phase in contact with fullers earth was difficult to control, due to the exothermic nature of the reaction. Further, the yield per ton of clay was very low.

Considerable study of the problem led to the conclusion that a process using high temperature and pressure and maintaining the distillate in the vapor phase would attain the desired results, be easily controlled, and prove more economical than other known methods of treating.

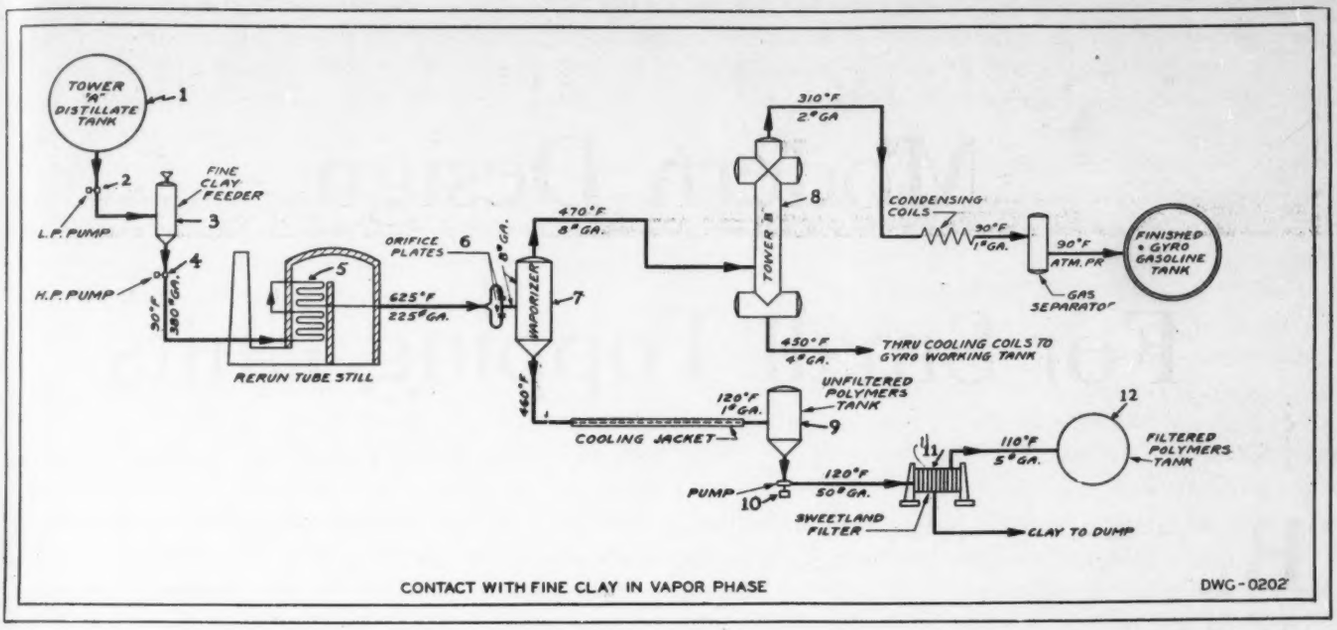
This process was designed primarily for the treatment of distillate produced by vapor phase cracking, but it is equally applicable to other cracked distillates.

It is well known that it is difficult to convert vapor phase cracked distillate into water white, stable gasoline without heavy losses in volume and anti-knock value. The highly colored compounds are very stable when compared with those found in liquid phase cracked distillates and they are present in higher concentrations. Vapor phase cracked distillates with a high percentage of olefines and diolefines is an excellent motor fuel if properly refined.

The activity of these unsaturated compounds is so great, however, that the removal of the colored constituents usually results in losing a substantial quantity of the desirable fractions. The problem has been, therefore, to devise a method which will selectively remove the undesirable colored compounds and those unsaturates which readily oxidize to form pseudo-gums, without the unnecessary destruction of the valuable motor fuel fractions. To a very large extent this has been achieved by this process.

DETAILS OF PROCESS

Numeral 1 indicates a storage tank for the raw distillate. This distillate is picked up by a low pressure pump (2) and discharged into a clay proportioned device (3) from which the mixture of distillate and clay is delivered by means of a high pressure pump (4) into the tube still (5). Under a pressure of 200 to 500 pounds, the distillate and clay pass at high velocities through the tube still where the temperature is raised to 600 to 650°F. The length of time at which the mixture is kept at this temperature, as well as the exact choice of temperature and pressure, is determined by the characteristics of the distillate under treatment, and the amount and kind of clay used and the quality of the finished product desired. Very fine fullers earth, bentonite, Death Valley clay and other products of this nature have been successfully used.



CONTACT WITH FINE CLAY IN VAPOR PHASE DWG-0202

Leaving the tube still, the mixture passes through orifice plates (6) to reduce the pressure. This method was found to be superior to the use of valves which were quickly cut out by the erosive action of the clay. The drop in pressure causes a drop in temperature and a vaporizing drum (7) is here provided to permit of separation of the clay and polymers from the gasoline fractions. The latter are delivered to a fractionating tower (8) which handles them in the conventional manner.

The mixture of polymers and clay is reduced in temperature to safe limits and passed to a tank (9) from which, by means of pump (10), it is fed into the filter press (11). The filtered polymers are stored in a tank (12) and the clay cake goes to the dump. The cake is fairly dry and the filtrate relatively clean oil. The use of all metal screens in this process has not proved entirely satisfactory, since it is very difficult to remove the finest particles of clay. The combination of metal screens and cloth blankets has been found to be satisfactory.

It is not known definitely why this operation should give so much better results on highly cracked gasoline than the usual method of vapor phase filtration through fullers earth or similar decolorizing agents. It is known, however, that this process affords maximum surface exposure between the refining agent and the oil under treatment and insures intimate contact of the oil vapors with the treating material wherein the refining properties of the clay are utilized to the fullest extent. It also permits of close control of the operating conditions. Temperature, pressure, time of contact and the percentage of water vapor are susceptible of very fine regulation. The oil vapors are not exposed to partially spent clay, which, in the percolation method of filtration results in undertreatment. Nor do the vapors come in contact with an excessive amount of fresh clay, producing an overtreated product. The exothermic nature of the reaction adds to the difficulty of temperature control in the ordinary vapor phase filtration process, but in this improved method it lends itself readily to the desired operation.

This process provides a continuously operating system wherein the oil to be treated is introduced at one end of the system and after treatment is continuously removed at the other end, and wherein the oil is subjected to fluctuating conditions of temperature and pressure, so regulated as to procure the maximum efficiency and to positively and effectively remove the undesirable materials from the treated oil. It also obviates the necessity of discontinuing the operation at intervals for renewing the clay.

The following data are representative of the vapor phase cracked distillate before and after treatment by this process:

	Raw distillate	Finished gasoline	Polymers
Percent	100	92.8	6.11
Gravity	55.3	55.1	18.7
F. B. P.	90	100	392
10%	148	158	508
20%	182	192	556
30%	212	222	588
40%	248	254	612
50%	284	288	632
60%	308	314	654
70%	348	340	679
80%	374	362	716
90%	400	386	760
Endpoint	424	410	
Color		30	
Gum		0.2 Mg	

Modern Design For Small Topping Units

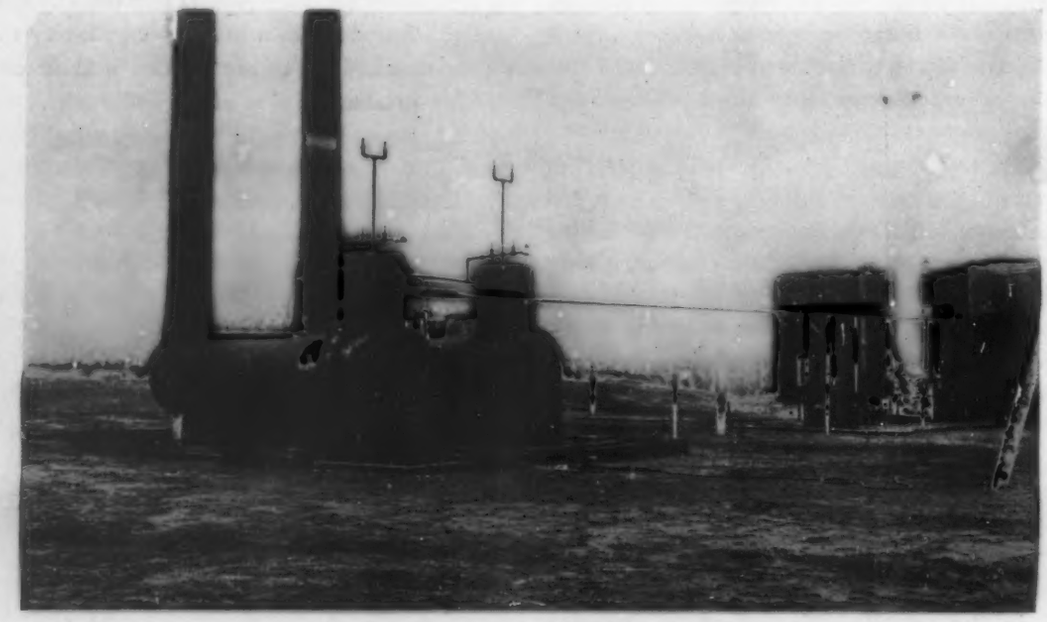
BY installing a small topping plant on his lease near Covington, Oklahoma, Charles E. Knox has succeeded in providing a market for both the crude oil and natural gasoline produced on this property. The distillate from the topping plant is blended with output of the natural gasoline plant and the gasoline sold through local channels. Fuel oil also goes into local consumption.

Having a small lease, whose output runs high in gravity, but at one time with an unsatisfactory outlet, Mr. Knox decided to erect a small unit to process this crude and blend the distillate with the natural gasoline manufactured on the lease.

Petroleum Engineering, Inc., in Tulsa, designed the plant. This company had foreseen the probable demand for small topping plants, and had designed equipment for that purpose. The design in this instance is based on a normal throughput of 200 barrels per 24 hours of 48° A. P. I. crude oil from which as high as 65 per cent may be taken overhead as a cut to be later fractionated into kerosene and gasoline. Should the future show the advisability of separating these two fractions, a second column may be installed for that purpose. As it stands, the plant is designed for continuous operation to take only one cut, at a time which is classed as a motor fuel whose end point is approximately 430°F.

Because of the steadiness of operation and the ease of control, the steam distillation type of still was selected; and since the steam requirements are small, the boiler installation was neither expensive nor elaborate. In this instance steam was desirable for other reasons. Gas from the wells is not abundant, while the fuel oil residuum from the unit was sufficient to provide fuel for the boilers. It also provides for the utilization of labor which need not be familiar with pipe still operation. Gasoline plant operators are as a rule more familiar with steam distillation than they are with fire distillation.

The equipment in the gasoline plant contains units which could be used for the topping plant as well. Tank-



Topping plant of Chas. E. Knox, located at Covington, Oklahoma

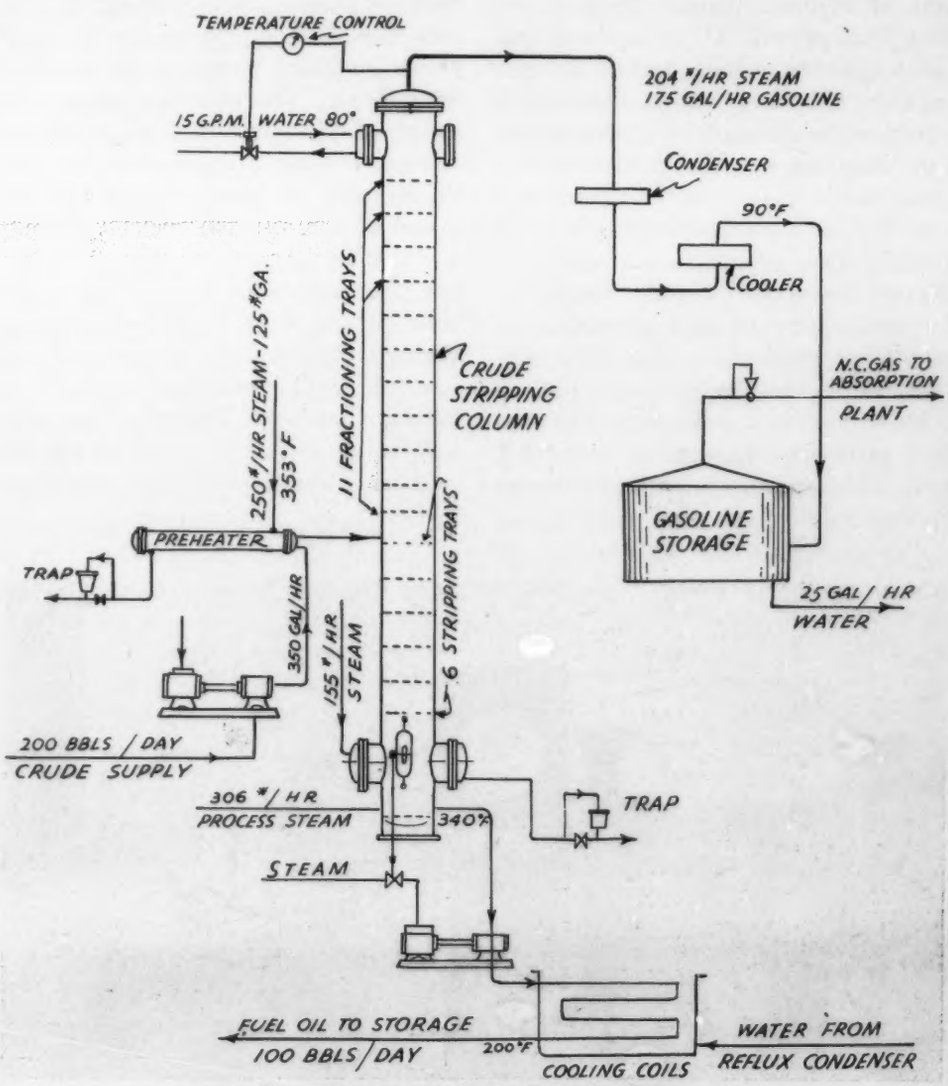
age for the casinghead condensate, naphtha tanks, water pumps for the cooling tower and the crude tankage on the lease were already installed. In order to tie in the new equipment so that it could be run in conjunction with the gasoline plant, or operated independently, only a small amount of auxiliary equipment was necessary. Oil field type boilers were on the lease as surplus equipment after drilling was completed. Two of them were available and were set up to furnish the necessary steam for operation and processing.

DESIGN AND OPERATION

The principal features of the installation are the simplicity of design and operation and the relatively small area necessary for its operation. That part of the topping plant which is surrounded by the auxiliary equipment is the combination still, fractionating column and reflux condenser. This unit is Petroleum Engineering, Inc., 30-inch diameter by 34 feet tall column having a reboiler or heating element incorporated in the base with six stripping plates directly above. The point of entry of the crude in process is just above the sixth tray while above this section are the 11 fractionating trays; all 17 are equipped with cast iron bubble caps. The entire column with its bubble caps and trays and the reflux condenser bundle at the top weighs only 8500 pounds. With this light weight, no elaborate nor expensive foundation was necessary; only a simple concrete monolithic slab was poured upon which to erect the column.

AUTOMATIC CONTROL

Since the crude is sweet, and after processing the distillate does not need chemical treatment the operation is a simple one and the designing engineers made every effort to incorporate simplicity of design into the unit. In operation, the unit is, perhaps, as nearly automatic as it can be designed for one man operation. The crude being received in the lease battery convenient to the plant, it is pumped from this tankage by a Fairbanks six by four by six-inch duplex steam pump. The rate of charge is governed by the velocity and pressure of the oil as it passes through an orifice placed in the



Flow Diagram of Crude Topping Plant

initial charging line. A line leads from the orifice flange to a steam governing valve placed on a by-pass around the throttle valve. As the size of the orifice was determined by the steam pressure at the boilers, the rate of throughput is constant and uniform. After passing through this orifice which is near the preheater, the oil enters the preheater to receive its first application of heat. Assuming a charge of 350 gallons per hour as the charging rate, and it may be determined by reading the manometer attached to the orifice flange, only 250 pounds per hour of steam at 125 pounds boiler pressure is required to obtain satisfactory temperatures on the oil as it passes through the preheater.

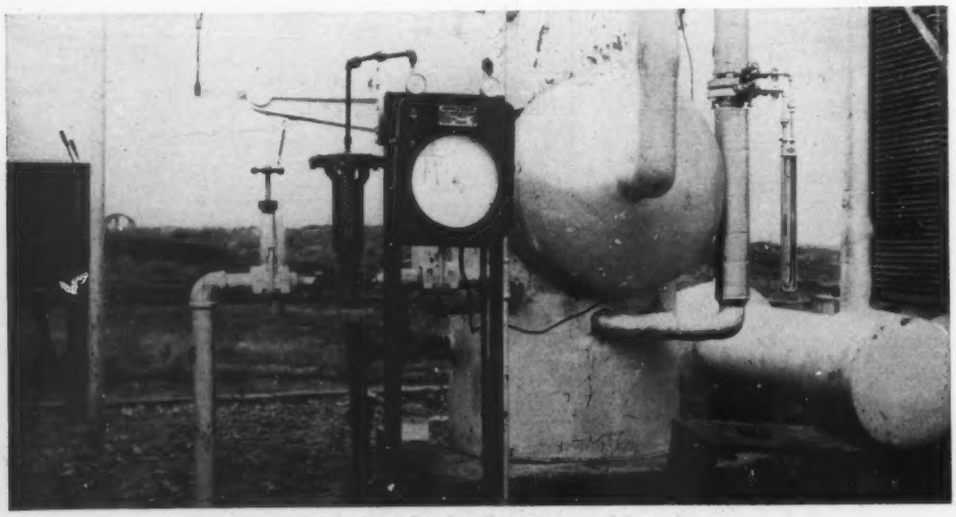
Entering the column, the oil during cold weather is carried at approximately 230 degrees. As it is introduced above the sixth tray of the stripping series, that which is not flashed directly to vapor is carried by gravitation downward to the compartment below these trays. The still heating bundle incorporated in the column requires approximately 155 pounds of steam hourly to raise the residuum to 320-340 degrees. Assisted by about 300 pounds as an hourly rate of process steam injected directly from the main steam header into the mass below the heating element, all of the remaining fractions not vaporized are driven from the oil. These vapors driven from the oil by heat from the reboiler bundle and the process steam, together with those vapors liberated at the point of introduction rise through the 11 fractionating plates above the stripping trays. Automatic reflux is continuous within the tower until, as the vapors pass the last tray near the top, reflux is completed by a water cooled bundle through which all the vapors must pass before being liberated from the column. Water is utilized for final reflux condensing, and approximately 15 gallons per minute are required to maintain a temperature of 280 degrees on the vapors leaving the top of the column. This temperature is automatically controlled by a Tycos recording temperature controller, operating a Tycos "Evenaction" one-inch diaphragm valve placed in the water line leading from the pump to the reflux section.

VAPOR RECOVERY

The vapors and uncondensed steam are carried to the cooling tower to be condensed in two sections of Griscom-Russell bentube atmospheric coolers. From these two sections, condensed gasoline and water are received in an accumulator or run tank for separation before being placed in storage.

In distilling an oil having a gravity as high as this, certain uncondensable gasses are liberated and pass through the plant under the two pounds carried on this unit. These vapors are removed from the run tank through a line leading to the intake of the gasoline plant compressor. A Clark 80 horse-power gas engine is utilized in the engine room, having tandem double acting compressor cylinders, one of which is 14 by 20-inch, and the other 7½ by 20-inch. Either plant may be operated in conjunction with the other, or operated independently.

As the system pressure is not sufficient to evacuate the fuel oil from the base of the tower, the residuum is removed by pumping. A Fairbanks 5¼ by 3½ by five-inch duplex steam pump was set beside the charging pump, connected directly to the base of the tower at the suction side. The rate of removal is controlled by the action of a Davis float box placed on the side of the column, through a lever connected to a steam valve in the line near the pump throttle. By this means, the removal of the stripped fuel oil is continuous and constant. When the control throttle is opened in the steam line, the pump starts automatically as the base fills up with oil and delivers it to two sections of American Radiator cast sections placed in the cooling tower to prepare the oil for storage or burning in the boiler combustion chambers. Fortunately in this case a satisfactory outlet for surplus fuel oil has been arranged for in some of the small towns in the neighborhood.



Equipment for control and preheating and piping arrangement at bottom of the tower. Chas. E. Knox topping plant at Covington, Oklahoma.

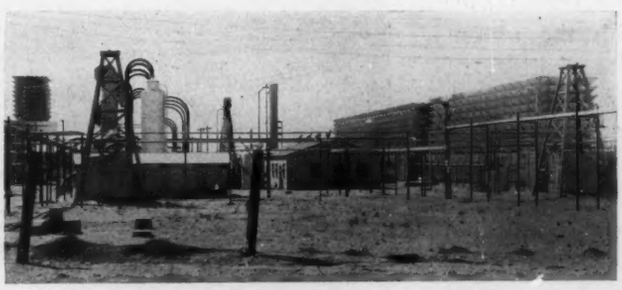
Coltexo Corporation's Lefors No. 3 Gasoline Plant

IN October, 1928, Coltexo Corporation began the construction of its Number 3 gasoline plant near Lefors, Texas, Panhandle. At that time there was available only enough gas to justify the installation of a 20,000-gallon Ereco absorption unit with the necessary distillation equipment. Only three type 80-165 horsepower Bessemer Cooper gas engines were placed on the foundations, and the gas passed through them required only half the number of absorbers usually set in that type plant.

Events moved rapidly from that time on, and during the ensuing year gas was developed with the oil production to such an extent that a greater expansion program was necessary. Placing the original equipment, as started, in operation on January 18, 1929, it was found within two months that two additional absorbers were necessary. Two months later, seven additional Cooper Bessemer "twins" were set on the foundations, and in August the second gasoline unit was erected. This new unit is a duplicate of the original installation, and is piped so that the gas can be passed to both at the same time, or from one to the other as necessity requires.

In October, 1929, just 12 months from the time the first ground was broken for the three engines, the plant was expanded so that the engine room contained 24 "twins". This showed an expansion of 800 per cent in power requirements, 200 per cent in distillation and 400 per cent in absorption equipment.

In September, 1931, under restricted production in that area, there were available only 54,000,000 cubic feet of gas for processing which contains an average of half gallon of gasoline. This is brought in under a vacuum of only three inches at the intake header. The forty-eighth 18 by 20-inch compressor cylinders boost this prssure at the discharge to 35 pounds through single stage compression. A battery of Griscom-Russell Bentube sections set in J. E. Prichard cooling towers bring the temperature of the gas down to an average of 70°F. before it is introduced into the eight Ereco absorbers. To prevent the entry of condensate, scrubbers were set near the absorbers. Lower temperatures are avoided to insure that the major portion of the gasoline in the gas enters the absorbers in the vapor phase. Higher temperatures are seldom encountered due to the usual high wind velocity and the low humidity of the atmosphere. Control of the lower temperatures is obtained by the application of varying amounts of water over the towers when unusual wind velocities are present, or when unusual low atmospheric temperatures prevail.



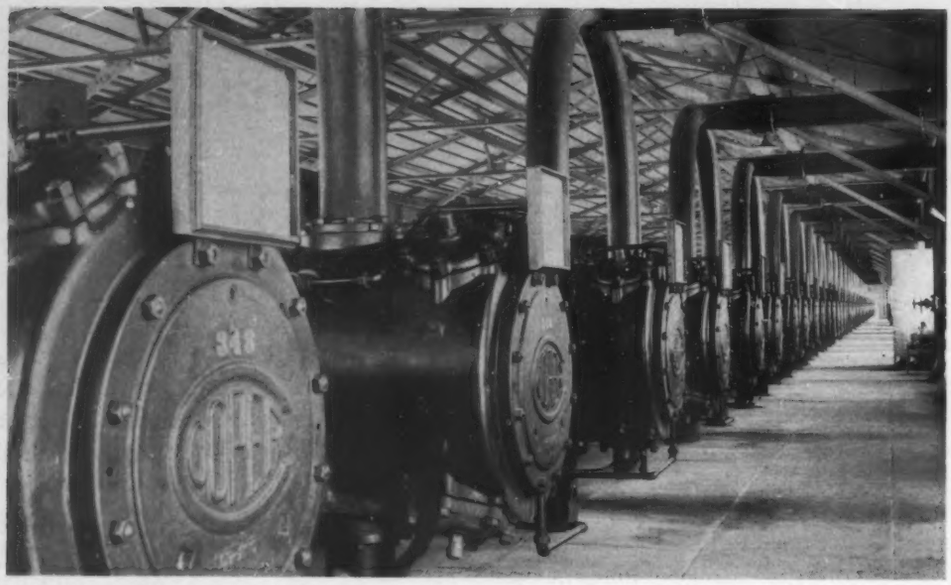
Looking west through the plant of Coltexo No. 3 near Le Fors, Texas

OIL-GAS RATIO CONTROL

Oil to the eight absorbers is moved by four, 14 by 12 by 12-inch Worthington duplex steam pumps placed in the room containing the other circulating pumps. A ratio of oil to gas is maintained at 35 gallons per thousand cubic feet; this amount being sufficient to strip all the desirable fractions from the gas. Under this operating procedure, the normal butane, practically all the iso-butane, and a large part of the propane, are extracted from the gas.

Control of the amount of oil to the individual absorber is effected by an unusual by-pass arrangement in the oil line leading to the absorber top. At a convenient height above the master gate on the oil riser, three two-inch lines were connected. The main line is blocked off by having a plate welded into it just below the point where the upper end of the two-inch lines are joined to the riser. More or less oil is admitted by opening as many of the three valves as is found necessary to maintain an equal division of the circulating mineral seal. Through this arrangement, no one absorber robs the others of its charge of absorption oil.

The rich oil leaving the absorbers is released by the utilization of Fisher float controls placed directly in the line, each absorber having its own controls. By-passes on these floats control settings are placed under the controls, with sufficient space between the two sets of fittings to enable the maintenance crew to clean them with a minimum inconvenience. It has been found



Compressor Room

that the more conveniently a piece of equipment is placed, the more likely it is to receive the proper attention. Oil to and from the absorbers is admitted and removed on one side while the gas is admitted and removed on the opposite side.

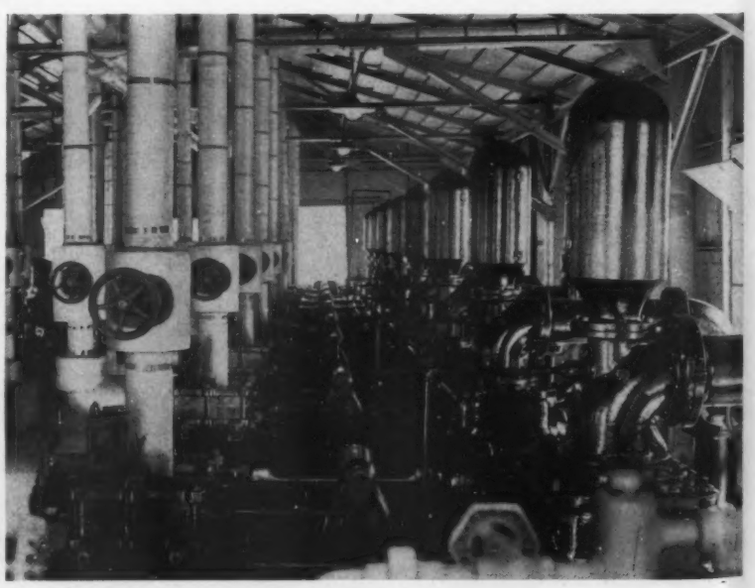
GASOLINE EXTRACTION

As the rich oil is removed, it flows into surge tanks from which it is picked up again by four pumps (the same size and kind as those used in circulating the lean oil) and passed through Griscom-Russell heat exchangers and preheaters to the evaporators. Steam with which to heat the oil prior to its entry into the evaporators, and steam for agitation and partial pressures in the evaporators is admitted to these units condensation free by the use of Armstrong traps placed near these units. A temperature of 345°F. is maintained on the two stills while the top of dephlegmators has a temperature of 230°F.

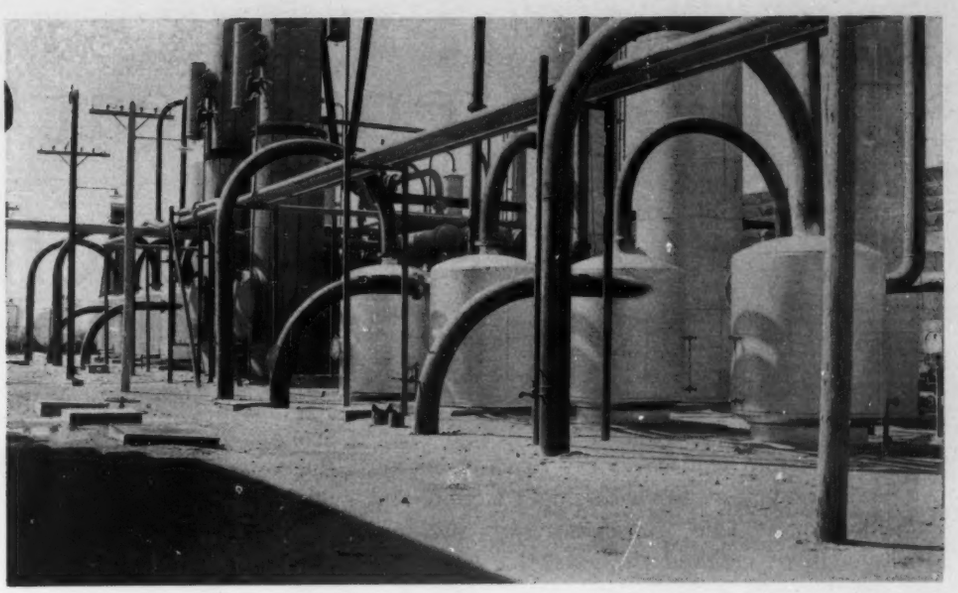
The gasoline vapors extracted from the rich oil by the evaporators are led through 10-inch lines to the cooling towers where they first pass through four, single-pass condensing sections to provide partial condensation. Only the more stable fractions are trapped out in an intermediate accumulator tank placed on the outlet of these condensers. This condensate is used as reflux when pumped back to the dephlegmator, the excess over that required passes out with the uncondensed vapors to the secondary battery of condenser sections. Re-entering the tower through three, three-pass sections of Bentubes, final condensation is effected, all of which is passed directly to storage to be stabilized.

Stabilization is carried out under 220 pounds with the Kettle temperature carried at 265°F. Reflux ratio of 10 to one is maintained on this particular gasoline to cut from it the propane and iso-butane. These lighter fractions after being separated from the charge are admitted to the residue line under fractionator pressure beyond the fuel take-off for plant consumption. This is arranged in this manner so that these lower boiling constituents of the gasoline may be transported to the carbon plant for conversion there. This plant is located a few miles east of the gasoline plant and consumes all the residue not needed for field plant consumption.

Water for plant use and consumption is produced in the yard from the characteristic Panhandle deep wells, three of which provide sufficient water for cooling the gas and oil and condensing the gasoline vapors and that used in the boiler room. It is lifted to the surface by the use of steam head operated pumps, the water from the tubing flowing directly



Oil Circulating Pumps



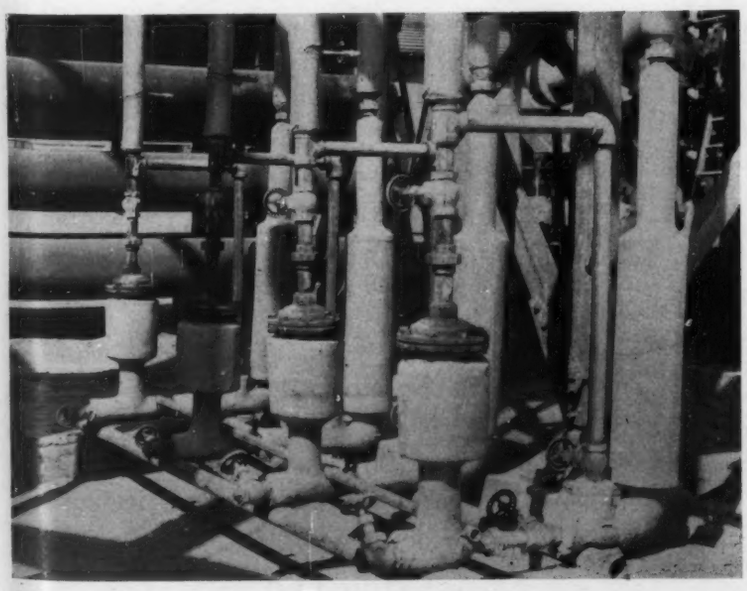
Piping Eliminates Sharp Turns

to the boiler supply tank and to the cooling pits under the louver towers. Circulation of water for cooling the gas and oil for condensation of the gasoline vapors is effected through the operation of three 12 by 14 by 12-inch Worthington low pressure duplex steam pumps placed in the row containing the three 12 by 18 by 12-inch Dean Brothers pumps which circulate the water over the engine jackets of the compressers.

ENGINE RECORDS KEPT

In the engine room, the 24, 165 h.p. type 80 Cooper-Bessemers are set in a row, with the intake coming in from the outside to join with a 160 degree pipe bend which is flanged directly on the connection plate of the compressor cylinders. The single line entering through the wall of the building serves each pair of cylinders, and is welded in the exact center of the 160 degree pipe fitting. Record cards are provided for each side of the individual engine, and are placed on the compressor cylinder head by being thumb tacked to a wood frame made especially for this purpose. The operators and repair men enter all the data on these record cards that pertain to the engine. All lubricating oil and repairs made are noted. Packing and ring installation are recorded as well as all other things done to the engine. Each one of the units has an individual company serial number and this number must be shown when any reference is made in reports or correspondence to the general or district office regarding the unit.

Other units in the plant have these serial numbers stenciled on them in a conspicuous place. All repairs to pumps or any other unit in the plant are charged directly to the unit. Any special packing or fittings placed in service is in this manner kept in recorded form and reference may be made to it at any time. Especially is this method of benefit when some particular packing or fitting is found that meets with the local need, and may be referred to again and again if found necessary.



Steam Trap Arrangement

Steam for operating the various pumps circulating water oil and for transporting the gasoline to the loading rack is generated by the use of 12 Walsh and Weidner Scotch marine type boilers. Six of the battery are 150 horsepower, while the other six in the boiler house are 100 horsepower. Water which is fed to these boilers is first softened in a Permutit water softener before being injected into them. It is soft when compared to water used in other fields, but scale-forming ingredients, in small amounts, are removed to prolong the hours on the line for each unit.

The Latent Heat of the Cracking Reaction

By S. N. OBRYADCHIKOV

THIS paper deals with the relation between the latent heat of the cracking reaction and such factors as the composition of the charging stock, yield of cracked gasoline, degree of cracking and temperature.

There are only three sources in the literature where data are given covering the latent heat of the cracking reaction:

(1) Leslie and Potthoff[1]
obtained this value experimentally by cracking paraffin wax with Al_2Cl_6 and found it to vary from 500-1000 calories per kilogram of cracked gasoline.

(2) Data for this reaction in the case of individual hydrocarbons have been compiled by Sachanen and Tilicheyev in their treatise (Liquid phase cracking).[2]

They give the latent heat of reaction for one kilogram of material, when totally converted, as equal to about 200 calories. For example:

C_6H_{14} $CH_4+C_5H_{10}$—280 (calories per kilogram)
$C_{10}H_{22}$ $C_5H_{12}+C_5H_{10}$—204 (calories per kilogram)
$C_{12}H_{26}$ $C_2H_4+C_{10}H_{22}$—188 (calories per kilogram)
$C_{10}H_{20}$ $2C^5H_{10}$—216 (calories per kilogram)
$C_6H_5C_4H_9$ $C_6H_5CH_3+C_3H_6$—209 (calories per klm.)

If we consider the total quantity of all cracked products (gases cracked gasoline and kerosene), then the amount of gasoline is less than 20 per cent of this total and we arrive at the figure of almost 500 calories per kilogram for cracked gasoline which is given by Leslie and Potthoff as the value for the latent heat.

(3) The author, together with Velikanov,[3] determined this value from the heat balance of Vickers cracking unit and obtained 350, 335 and 365 calories per kilogram of cracked gasoline.

Since it was very important to have more reliable data for thermal calculations, the author therefore undertook this task and has reported more than 50 results of tests on the value of the latent heat.[4]

[1]Ind. Eng. Chem.: 18, 776-85 (1926); C. A.: 21, 169 (1927).
[2]Translation of which is soon to be published.
[3]Neftyanoye Khozyaistvo: 17, 370.2 (1929); C. A.: 24, 1206 (1930).
[4]S. N. Obryadchikov. The heat problems in the cracking process. Neftyanoye Khozyaistvo: 1931, special number.
[5]A. N. Sachanen. Summary of Investigation of Grozny Petroleum Oils. 1927, Moscow.

METHOD

The latent heat of reaction can be calculated from the following equation:

$$Q\text{react.} = \Sigma\, Q\text{in.} - \Sigma\, Q\text{fin.}$$

where

Qreact. = the latent heat of reaction,
ΣQin. = the total heat of combustion for the initial products,
ΣQfin. = the total heat of combustion for the final products of the reaction (since the contents of hydrogen and carbon are constant before and after the reaction).

The following equation was used for calculating heat of combustion for the liquid products (Cragoe. Misc. Publ. No. 97, Bureau of Standards, "Thermal Properties of Petroleum Products").

$$Q_v = 12400 - 2100\ d^2$$
$$Q_p = Q_v - \%H \cdot 50.45$$

where

Q_v = total heat of combustion at constant volume (calories per kilogram),
Q_p = lowest heat of combustion at constant pressure (calories per kilogram),
d = specific gravity at 15°C.
$\%H$ = per cent of hydrogen in the products, determined by the following relation: $\%H = 26 - 15d$.

This equation gives the same results (the difference is less than one per cent) as the experimental data obtained for Grozny petroleum oils.[5]

Let us assume 7840 calories per kilogram as the value of the heat of combustion for coke and 11400 calories as the value of the heat of combustion for cracked gases. The latter value was calculated from analysis of the cracked gases from the Vickers unit.

In order to check this method, the results of steam distillation without any cracking were calculated and found to be as follows:

$$\Sigma Q\text{in.} = \Sigma Q\text{fin.}$$

One of these calculations is given in table 1.

TABLE 1.

Steam distillation of the bottoms from crude containing little paraffin, (Well No. 5/233, crude, Old Grozny field, salt ravine).

d_{15} 0.909; $Q_p = 100045$ cal. per kg.

Nos. of fractions	% of bottoms	d_{15}	Q^p cal. per kg. of fraction	Q^p cal. per kg. of bottoms
1	8	0.839	10235	818.8
2	8	0.846	10210	816.8
3	8	0.856	10195	815.6
4	8	0.867	10160	812.8
5	8	0.879	10130	810.4
6	8	0.890	10105	808.4
7	8	0.905	10055	804.4
8	8	0.910	10040	803.2
Residues	36	0.950	9915	3569.4
	100			10059.8

Difference $\Delta = 10059.8 - 10045 = 14.8$ cal.

The average difference between ΣQin. and ΣQfin. is about 15 calories which is the average error we may have made, and this error will amount to about 15 per cent when we assume that this difference is about 100 calories. The total error including a possible inaccuracy in the amounts of gases and other products as well as in the specific gravity will be about 20-25 per cent. This error is possible in the data given below.

The experimental data given in the above mentioned book of Sachanen and Tilicheyev were taken as a basis for this calculation, the numbers of the experiments corresponding to those in this book. Two calculations in Table 2 may be cited as example.

TABLE 2.

Number of experiment Charge	12 Grozny paraffin distillate		29 Grozny parafin residue	
d_{15}	0.883		0.906	
Fraction below 200°C	0.3		....	
Fraction below 300°C	3.0		4.8	
Conditions of test				
Temperature °C	450		425	
Pressure atmospheres	12		10	
Duration minutes	8		62	
The cracked products yielded in % by weight				
Gas and loss	4.9		5.6	
Fraction below 200°C	18		17.8	
Fraction below 200-300°C	23		17.7	
Residue	52.5		57.1	
Coke	0.1		0.9	
Amount of cracked products	98.5		99.1	
Specific gravity				
Fraction below 200°C	0.741		0.733	
Fraction below 200-300°	0.832		0.836	
Residue	0.911		0.933	
Heat of combustion				
Gases	11400	.0.049=558	11400 . 0.056=	639
Fraction below 200°C	10485	.0.18 =1887	10500 . 0.178=	1869
Fraction below 200-300°C	10260	.0.23 =2360	10275 . 0.177=	1819
Residues	10035	.0.525=5268	9970 . 0.571=	5693
Coke	7870	.0.001= 8	7870 . 0.009=	71
Heat of combustion	10058		10091	
Heat of combustion of the charge	10125	.0.985=9973	10055 . 0.991=	9965
Latent heat of reaction				
Calories per kg. of crude	85		126	
Calories per kg. of cracked gasoline	472		708	

RESULTS AND CONCLUSIONS

A. Effect of the composition of the charge.

The following example may be cited as an illustration:

We may have two hydrocarbons, (1) R-R and (2) R-R-R, where R are radicals the size of a gasoline molecule. If we intend to produce one kilogram of gasoline from one kilogram of hydrocarbon (1) we must supply heat to sever the bond between R-R. More heat must be supplied, of course, to break two bonds as is the case with hydrocarbon R-R-R (2). This leads to the conclusion that cracking of heavier oils requires more heat than cracking of light oils. The calculation given in Table 3 confirms this assumption.

TABLE 3.

Charging stock	Specific gravity	No. of the experiment	Yield per cent of cracked gasoline	Latent heat of cracking reaction in calories/kg. of cracked gasoline
Gas oil	0.833	1	12.5	—344
Grozny		2	16.7	—540
paraffin		3	29.7	—270
distillate	0.883	7	19.6	—650
		8	14.0	—536
		9	21.5	—569
		10	34.2	—407
		12	18.0	—472
Grozny		13	28.9	—574
residuum		14	34.0	—676
paraffin base	0.906	17	11.2	—830
		28	11.3	—868
Surakhanui		29	17.8	—708
residuum		32	9.8	—722
Baku	0.890	43	9.4	—680
		45	23.7	—536

B. The relation between heat of reaction and yield of cracked gasoline.

Because of the fact that gasoline obtained as the result of cracking of all products is taken into consideration, it is evident that it will be more convenient to calculate the latent heat of the reaction in calories per kilogram of cracked gasoline, and the latent heats of formation of all other products are based on the latent heat of the cracked gasoline.

If the reaction proceeds mostly with the formation of intermediate products and with very low yield of gasoline, then the latent heat value of the reaction must be very high.

At the beginning of the cracking process we have such a reaction condition and therefore at first very high values for latent heats of reaction (See Table 4).

TABLE 4.

Charge	Number of the experiment	Per cent of cracked gasoline	Latent heat of the cracking reaction in cal./kg. of cracked gasoline
Grozny paraffin			
distillate	5	4.2	—1780
	6	9.8	— 898
	7	19.6	— 650
Grozny res-	9	21.5	— 569
iduum (par-	14	34.0	— 676
affin base)	16	5.5	—1235
	25	6.5	— 982
	26	7.8	— 897
	29	17.8	— 708

In Table 5 are given the average values of the

latent heat of the cracking reaction based on the cracked gasoline yield; these values were obtained by interpolation of the results from 30 calculations.

TABLE 5.

Yield of cracked gasoline per cent	Latent heat of reaction in calories per kilogram of cracked gasoline: Average from the beginning	Approximate at the given yield
5	—1500	—1000
10	— 950	— 700
15	— 650	— 500
20	— 500	— 380
25	— 480	— 300
35	— 440	— 200

In the second column are shown the average data for the latent heats from the beginning of the reaction to the given yield of cracked gasoline; and in the third column—approximate values of the latent heat at the given yield of cracked gasoline.

EXAMPLES

1. What quantity of heat would be required to produce 15 per cent of gasoline from 100 kg. of charging stock?

$$100 \cdot 0.15 \cdot 659 = 9750 \text{ calories.}$$

2. What quantity of heat would be required to produce five per cent of gasoline from 100 kg. of charging stock if 20 per cent were already formed?

$$100 \cdot 0.05 \cdot \frac{(380 + 300)}{2} = 1700 \text{ calories}$$

[6]Neftyanoye Khozyaistvo: 16, 223-9 (1929); C. A.: 23, 4807 (1929).
[7]Oil and Gas J.: March 6, 1930, p. 290.

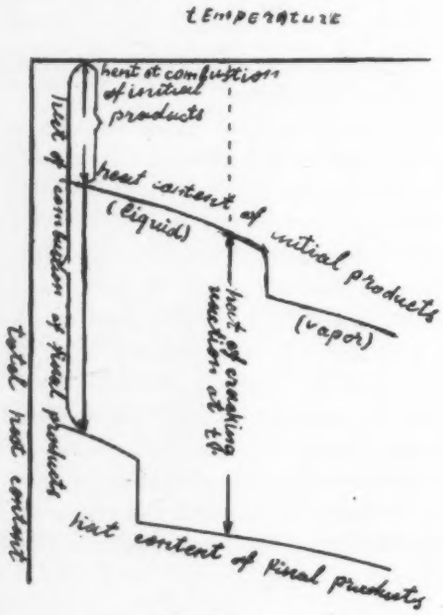

temperature
heat of combustion of initial products
heat content of initial products (liquid)
heat content of final products (vapor)
heat of cracking reaction at t°
total heat content
heat content of final products

FIGURE 1

C. THE INFLUENCE OF THE DEGREE OF CRACKING

There are two types of reactions in the cracking process

1. Cracking reactions with the absorption of heat (endothermic);

2. Polymerization and condensation reactions with the liberation of heat (exothermic).

The first named reactions occur at the beginning of the process, but the second have a great importance at the end. It is possible that by heavy cracking there will be a certain quantity of heat given off near the end of the process.

In Table Number 6 are given values for heavy cracking. The experimental data for the calculation are taken from an article by Sachanen and Tilicheyev.[6]

TABLE 6.
Cracking of Gas Oil (Emba Crude)

Type of cracking	Specific gravity of cracked gasoline	Per cent cracked gasoline	Latent heat of reaction cals./kg. of cracked gasoline
Primary	0.746	28.7	—477
Secondary	0.769	21.3	—400
Tertiary	0.797	16.3	—178
Quaternary	0.833	12.1	+388

D. INFLUENCE OF TEMPERATURE

This factor can be determined from the following equation G. Burrell.[7]

$$\Delta Q_{reac.} = \int_0^t \Delta \cdot dt$$

Where:

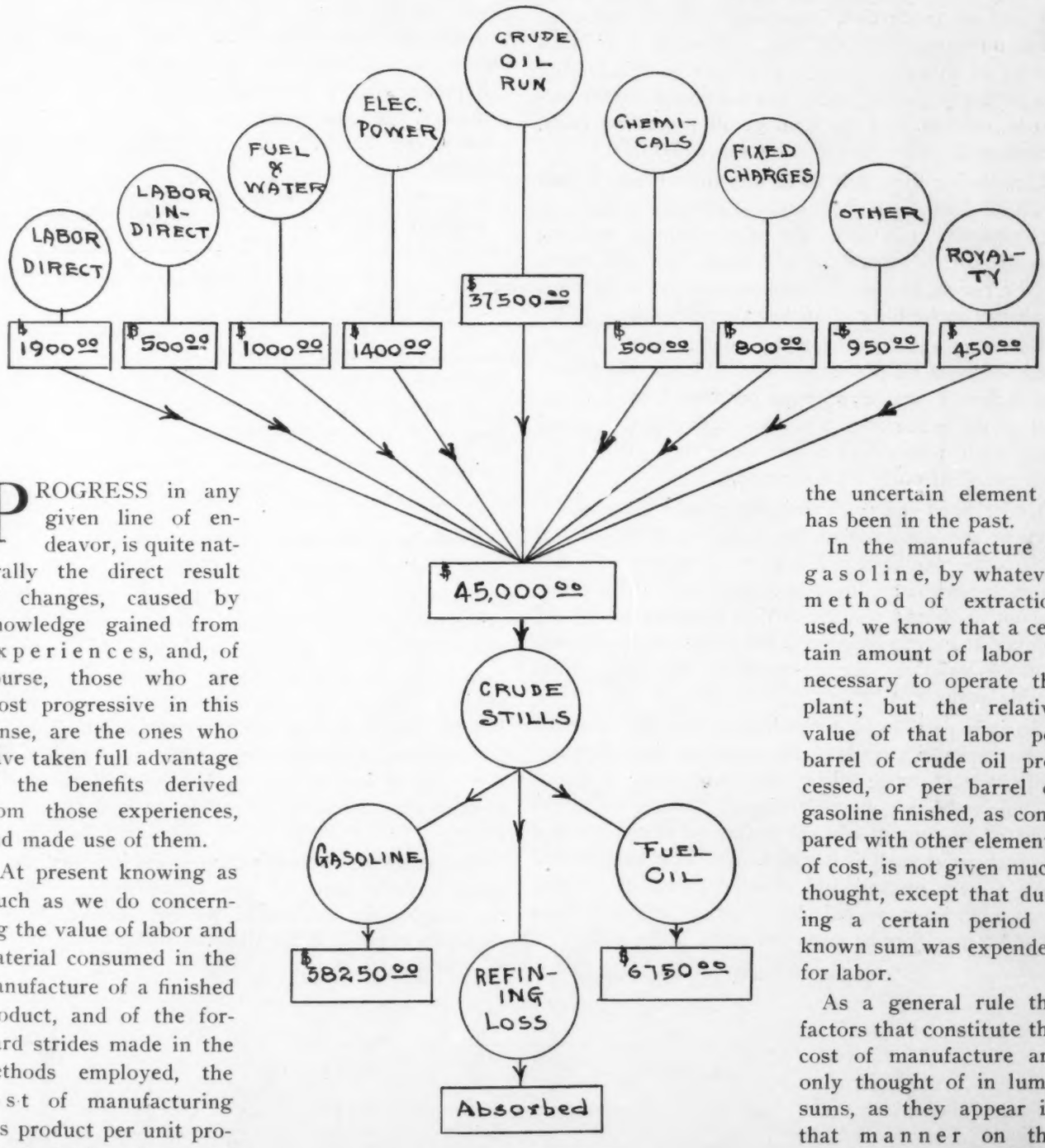
$\Delta Q_{reac.}$ is the difference between the values of the latent heat of reaction at the temperature 0 and t.

ΔC is the difference between the values of the heat contents at the same temperatures.

The graph shown in Figure 1 indicates the change of the latent heat of the cracking reaction with the temperature is not high and can be disregarded.

Elements in Refining Costs

By R. E. KERFOOT



PROGRESS in any given line of endeavor, is quite naturally the direct result of changes, caused by knowledge gained from experiences, and, of course, those who are most progressive in this sense, are the ones who have taken full advantage of the benefits derived from those experiences, and made use of them.

At present knowing as much as we do concerning the value of labor and material consumed in the manufacture of a finished product, and of the forward strides made in the methods employed, the cost of manufacturing this product per unit produced, need no longer be the uncertain element it has been in the past.

In the manufacture of gasoline, by whatever method of extraction used, we know that a certain amount of labor is necessary to operate the plant; but the relative value of that labor per barrel of crude oil processed, or per barrel of gasoline finished, as compared with other elements of cost, is not given much thought, except that during a certain period a known sum was expended for labor.

As a general rule the factors that constitute the cost of manufacture are only thought of in lump sums, as they appear in that manner on the monthly cost and yield re-

port. In the use of chemicals, we know from practice, that it requires a certain number of pounds of sulfuric acid to produce a required condition in the processing, and that we can readily determine if so inclined, the value of the acid consumed.

But what of the many other factors? How much fuel consumed? What per cent of the total cost was the value of fuel consumed? How much paid for electric power purchased? And, how much for many other things? All of these facts may be easily determined and for the purpose of illustration, the chart included, has been constructed. The complete detail is entirely contained in Table 1.

Here, it wil be seen, are all of the items included in the cost of production, beginning with the value of crude oil consumed. The cost, delivered in refiner's storage of this crude oil, is $1.25 per barrel. During this particular period, there was consumed 30,000 barrels in refining, and the total of all processing costs, including the value of crude oil was $45,000.00.

Considering then, that there was run to stills a total of 30,000 barrels of crude oil, and that the total of all expenses was $45,000.00, the other elements entering into the cost of operation, i.e., labor, fuel and water, electric power, etc., required an expenditure of 25 cents per barrel above the cost of the raw crude oil.

Referring again to the table, we find that for each barrel of crude oil processed, the total labor cost (direct and indirect), was eight cents per barrel, or 5.33 per cent of the entire cost. Also that the amount required for electric power was nearly five cents (.0467) for each barrel of crude oil consumed.

Under fixed charges, consisting mainly of taxes, insurance and depreciation, the cost for this period is .0266 cents per barrel. However, as expenses in this group are calculated upon an annual basis, the monthly amount per barrel will fluctuate in accordance with the number of barrels processed; the greater the consumption, the lesser the amount applicable to fixed charges in this analysis.

As it is possible to determine, in advance of actual performance in a refinery, the expected yield of a certain grade of crude oil by laboratory tests; it is also possible to know quite accurately in advance every element of processing cost, providing of course, that a comprehensive analysis is maintained. This analysis will have a very marked and material effect in creating greater efficiency, as, only by knowing individual items of costs, may they be improved upon.

TABLE 1
ELEMENTS IN REFINING COSTS

Consumed—	Amount	Per Bbl.	Per cent of Total
Crude oil consumed (30,000 Bbls.)	$37,500	$1.25	83.34
Direct labor; Firemen, stillmen, operators	1,900	.0633	4.22
Indirect labor; Helpers, pumpers, laborers..	500	.0167	1.11
Fuel and water	1,000	.0333	2.22
Electric power	1,400	.0467	3.11
Chemicals used	500	.0167	1.11
Fixed charges; taxes, insurance, depreciation	800	.0266	1.78
Others; Superintendence, office, laboratory, general plant repairs, etc.	950	.0317	2.11
Royalty; Account of license to use a patented process	450	.0150	1.00
	$45,000	$1.50	100.00

Produced—	Yield Per cent	Barrels			
Gasoline	65	19,500	$38,250	$1.9627	85.
Fuel Oil	30	9,000	6,750	.75	15.
Loss	5	1,500			
	100	30,000	$45,000		100

In the cost and yield section of the table of operations, fuel oil produced is valued at the current market price of 75 cents per barrel. Loss in refining is not evaluated, as it has been considered as an intangible part of the cost of gasoline produced. This method is used here; for it is the simpler one, and as the entire subject is treated in only a general way, we consider it sufficiently consistent for the purpose intended.

There are, of course, other methods used to determine the value of loss in refining, but we shall not go into their merits; for some are rather involved and complicated.

The value of cost analysis and cost information, depends entirely upon the uses to which it is put; and the simpler the method is, without sacrifice of vital facts, the better it is for all concerned.

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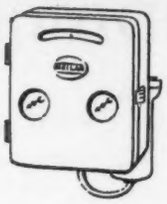

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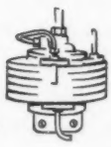

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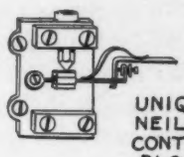

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Calculating the cost of the time taken from other work, coupled with the danger of burns from coming into contact with leaking drums, the management of Root Refining Company decided to lower expense and danger both at the same time. Receiving this treating material in carload lots and using it in rather large quantities, the expense of lifting the drums to the steaming tanks ran into considerable money. Each morning it was necessary for the yard foreman to detail a few men to assist the light oil treater in handling the day's requirements.

A shed was built that will protect a carload of drums. Under it a unique series of supports were constructed. Utilizing four-inch discarded pipe for piers, cross braced for rigidity, and cross-members of pipe welded to the top of these piers, runways of angle iron were laid to support the drums. These angles were laid back to back so that four rows of drums are in the shed at one time. One end, that farthest from the steaming tanks, was built just high enough that the trucks handling the caustic could back up to it with the tail board level with the angle iron storage runways.

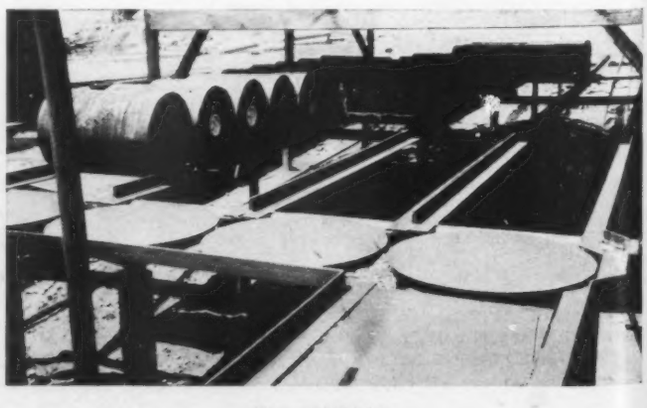
The other end of the rack, from which the drums are moved to the steaming racks, was built quite close to the melting tanks. Then the problem of transferring the drums from the three runways to the one in line with the melting tanks was solved. At the end of all four runways a turn-table was built so the treater could handle the drums without lifting, and without being forced to call for help or to use a chain hoist in moving them. Setting six-inch pipe in the ground for the turn table supports, cementing them in for rigidity and to maintain alignment, the tables were placed in the upper end of the supports.

To make the tables, heavy boiler plate was used by cutting circular sheets large enough in diameter to accommodate a drum and with room to spare. These were welded to a piece of four-inch pipe, bracing the sheets with triangular pieces of the same metal as the tables by welding the braces to both the plate and four-inch pipe. A ring was formed and welded to the outer edge of the tables and to the outer tips of the brace angles. The lower part of the four-inch pipe was flat headed so that a large ball, which was placed in the six-inch supports, could be used as a support and to provide a method of turning the table with ease with a drum of caustic on it.

A fender of angle iron was provided near the tables to prevent a drum rolling from them as the treater moved them from the storage rack to the steaming tanks. When a drum is needed from the runway farthest away, it is rolled to the table at the end of that section. As it is placed on the table, it is given a shove turning it at right angles to its original position on the rack. By rolling across the other tables, it comes to rest on the table in line with the runway leading to the tanks.

Give it another twist and it is in line for immediate use with no lifting, the drums are not rotted by being exposed to the elements. No danger of dropping the drums exists, because every precaution has been taken to direct the drum in its travel from storage to the point of steaming by these angle iron supports and guides. It was estimated that by using material from unused supply, and by salvaging a part from the old pipe piling, the apparatus did not cost to exceed $350.00

Detail of Table Supports



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